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-ECAS-
WESTINGHOUSE PHASE I FINAL REPORT
Volume VIII —OPEN-CYCLE MHD

by

D.Q. Hoover, et al

WESTINGHOUSE ELECTRIC CORPORATION RESEARCH LABORATORIES

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16. Abstract

Plant costs and efficiencies are presented for three basic open-cycle MHD systems: direct coal fired system, a system with a separately fired air heater, and a system burning low-Btu gas from an integrated gasifier. Power plant designs are developed corresponding to the basic cases with variation of major parameters for which major system components are sized and costed. Flow diagrams describing each design are presented. A discussion of the limitations of each design is made within the framework of the assumptions made.

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- J. A. Dilmore, who made the combustion calculations necessary to define the required preheat temperatures for the main combustor and the separately fired air heater, and did the combustor design and pricing.
- R. Draper, who did the work on the steam generator design and pricing.
- P. W. Eckels, who did the direct and separately fired air preheater design and pricing.
- E. J. Lahoda, who described the necessary subsystem to remove sulfur from the stack gas by chemical conversion of seed material. This included the detailed description, sizing and costing of the subsystems many components and the calculation of the needed power for its operation.
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- R. E. Voshall, who prepared the size, weight, and costs of dc to ac power conditioning system for open-cycle MHD generators.
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SUMMARY

The open-cycle MHD study includes three basic MHD systems each of which is bottomed by a nearly conventional 24.132 MPa/811°K/811°K (3500 psi/1000°F/1000°F) steam turbine generator. These systems are a direct fired coal burning system using the potassium seed for sulfur removal, a carbonized coal (char) burning system with a direct fired air preheater, and a low-Btu gas fired system using cesium as seed and utilizing an integrated low-Btu gasifier with in-bed sulfur removal.

The direct coal fired system represents the simplest of the plant designs. This plant shows the lowest overall cost of electricity [ranging around 7.78 mills/MJ (28 mills/kWh)] and an overall energy efficiency of about 47%. The power requirements of the auxiliaries and the seed treatment plant of this system were high, decreasing plant efficiency approximately 3 points. The seed (potassium carbonate) is partially regenerated and elemental sulfur is recovered as a by-product.

Various parametric cases are also studied for each of these systems. These represent variations in coal type, coal moisture content, ash carryover, air preheat temperature, and power plant size. Except for a few instances, the net impact of varying these parameters on the cost and performance of the "revised" plant design was relatively small (probably within the uncertainty of the analysis). The exceptions involved certain high ash carryover situations and cases where large pressure ratio changes occurred (where duct size and consequently magnet costs were affected).

The system with the integrated low-Btu gasifier shows substantially higher capital cost than either of the other cases. Three factors are involved: (1) cost for the gasifier, (2) a significant cost increase in the magnet due largely to the higher MHD pressure ratio that was used,

and (3) increased hydroxyl ion in the gas accumulating from the gasification process. Use of the higher pressure ratio was an attempt to minimize gasifier cost; however, the associated increased magnet cost was found to more than compensate for this saving. Lower system pressure ratios are suggested. The cost of electricity for this system ranges around 10.55 mills/MJ (38 mills/kWh) with an overall energy efficiency of 49%.

The system burning carbonized coal in the combustor uses the gapor from the carbonizer to fire the air preheater. As might be expected, this plant design showed higher costs than the direct-fired MHD case, due mainly to the additional costs of the separate air preheaters and carbonizérs. This plant has a cost of electricity of 9.17 mills/MJ (33 mills/kWh) with an overall energy efficiency of about 48%.

The general insensitivity of cost and performance to the parametric analysis does not mean that the individual parametric variations did not influence the system significantly. What appeared to occur was a series of compensating effects such that the overall cost was not greatly altered. For example, changing fuels from the bituminous (10% moisture, 3.9% sulfur) to the lignite (27% moisture, 0.7% sulfur) decreased combustor temperature. To compensate, the design required increasing channel size, thereby, increasing magnet costs. This cost, however, was offset somewhat by savings in seed-treatment because of the lower sulfur content of the lignite. This insensitivity of MHD performance and cost to major parametric variations suggests that each of these basic system designs have some flexibility to adapting to different and changing utility applications.

The open-cycle MHD topper with its high efficiency does have the potential for a future base load power system. A final judgement on the commerical viability of this system will require establishing better estimates for the cost of superconducting magnets, recovery heat exchangers and the air preheaters. These items represent approximately 40% of the total direct-system cost. Demonstrating viable, large-scale MHD channel and combustor designs should be given high priority in MHD research programs.

9. OPEN-CYCLE MAGNETOHYDRODYNAMICS

9.1 State of the Art

Open-cycle MHD is a developing technology which will require considerably more research and development before it is reduced to commercial practice.

Recent accomplishments include one-hour continuous operation of a generating duct with direct coal firing (Reference 9.1). No mention is made of the duct operating temperature or of the plasma conductivity, which are critically important to a successful commercial operation. In this facility the coal was burned with oxygen, eliminating the need for an air preheater (Reference 9.2).

Considerably longer-duration runs are reported in Reference 9.3. Runs of more than 10 hours at full power have been made. The fuel used in this facility was a light fuel oil, and the oxidant was oxygen-enriched air. Gas conductivities of 10 to 12 mhos/m are reported.

From the above references, it seems that considerable progress in the design and construction of small MHD generators has been made. It is reasonably certain that a large-scale MHD generator could be successfully designed and built using either electrode replenishment (as advocated in Reference 9.2) or cooled walls (Reference 9.3), although all the problems have not yet been resolved.

Superconducting materials and the art of designing large magnets has progressed to the point where there appears to be little doubt that the necessary magnet can be designed and built successfully. The cost of these magnets, however, is not well known. Estimates used in economic calculations (including this report) assume sharp reductions in the cost of superconducting wire, which may or may not materialize. Solid-state conversion of the MHD output also appears to be well in hand.

A major problem area remaining is the development of suitable heat exchangers. In order to obtain attractive efficiencies, it is necessary that the heat in the exhaust products be recovered and that a substantial amount of this heat be transferred to a high-temperature [1589°K (2400°F)] fluid to limit its thermodynamic degradation. The most comprehensive study of these problems was sponsored by the Central Electricity Generating Board of Great Britain (Reference 9.4). The problems were corrosion and deposition on the heat exchanger surfaces due to the seed material and any ash which was in the stream. The CEGB study did not find solutions to these problems, and it may be significant that their work on open-cycle MHD has been at a much-reduced level in recent years.

It appears that some presently popular solutions to the duct problem (replenishment of electrodes by slag or ceramics injected to coat the electrodes) will aggravate the heat exchanger problem. In Reference 9.3 it is claimed that the slag will be separated from the stream in a cyclone at temperatures above 2000°K (3140°F). Separation of slag from the gas stream at this temperature will prove to be extremely difficult since most of it will still be liquid and some (depending on the coal) vapor at this temperature. Any slag which is carried over will tend to form a very tenacious coating on the cooler heat exchanger surfaces (Reference 9.4). This problem may be eliminated by the schemes assumed in this study (Appendices A 9.2 and A 9.3), but this has yet to be demonstrated.

Another area of uncertainty is the recovery of seed in a form suitable for removal of sulfur from the gas stream. There is a lack of fundamental data on the chemical processes involved. Appendix A 9.1 discusses this problem in considerable detail and describes a system which should work but whose equipment sizes and costs are uncertain. The capital cost of the system is substantially below that reported for scrubber systems, but its power and energy requirements are rather high for the bituminous (3.9% sulfur) coal. Since the energy, power, and capital costs are directly proportional to the quantity of sulfur removed (rather than to the volume of gas handled), this system is attractive where less sulfur

is to be removed (for example, the subbituminous and lignite coals of this study).

The potential performance of open-cycle MHD is very attractive. With the cycle conditions envisaged, overall plant efficiencies approach 50%. Since the MHD generator is essentially a volumetric device, its economics improve with plant size, and no fundamental limitations are now known. Emissions of sulfur can probably be controlled by the seed material at some cost in energy and capital (Appendix A 9.1). The emissions of nitrogen oxide can also be controlled by eliminating any excess oxygen in the high-temperature regions of the cycle and injecting air to complete combustion in some low-temperature regions where insignificant quantities will form. If the emissions of seed material are treated simply as particulate matter (as assumed in this study), recovery efficiencies of about 99.5% are sufficient to satisfy current EPA standards.

9.2 Description of Parametric Points to Be Investigated

For open-cycle MHD there are three basic cycles with variations of parameters for each basic cycle. The basic cycles and their variations are discussed in order of increasing complexity rather than in numerical order.

9.2.1 Base Case 2

This cycle is the simplest of the three basic cases. As shown in the schematic diagram (Figure 9.1), the prepared coal is fed directly to a single-stage combustor, where it burns with preheated combustion air. To limit the formation of nitric oxides, the combustor is operated rich (0.95 times stoichiometric air). The maximum temperature of the products of combustion is a function of the fuel (bituminous coal as received), coal pretreatment, stoichiometric air-fuel ratio, combustor heat loss, and the air preheat temperature [$\sim$ 1589°K (2400°F)] at the specified combustor pressure.

Five percent of the heating value of the coal fired was assumed to be transferred to the wall-cooling fluid or rejected with the molten slag. A heat loss to the combustor walls and slag tap of 5% of the fuel

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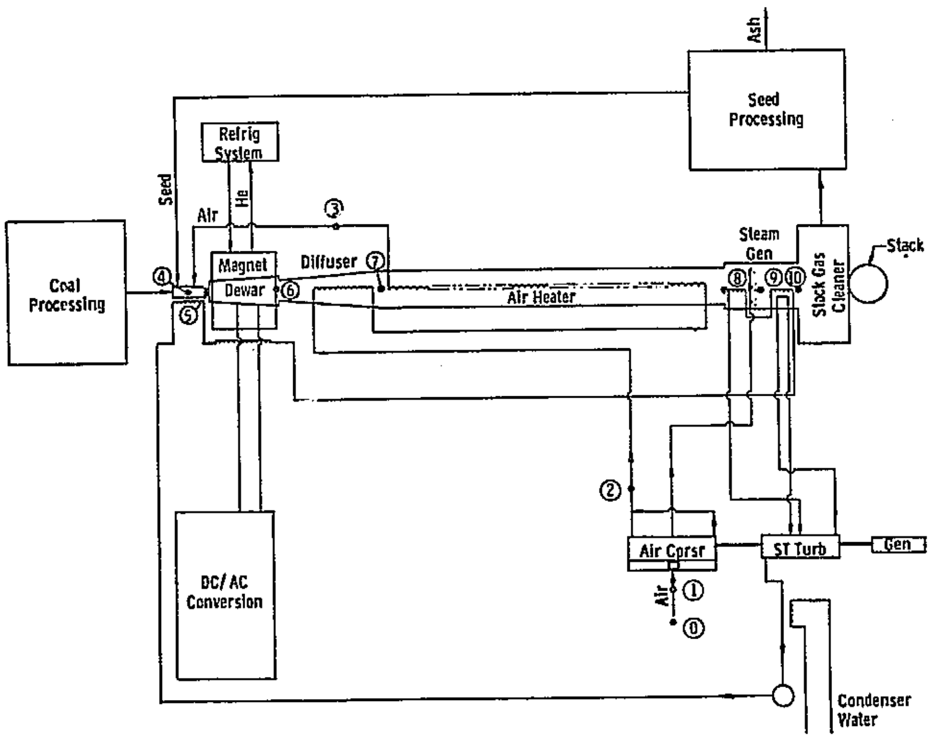

Location	Point No.	Pressure, Psia	Temperature, °F	Flow, lb/s
Ambient	0	14.696	59.0	2768.5
Compressor Inlet	1	14.40	59.0	2768.5
Compressor Outlet	2	95.36	465.2	2768.5
Preheater Outlet	3	92.58	2398.4	2768.5
Combustor Outlet	4	88.18	4414.4	3144.3
MHD Duct Entrance	5	59.18	4185.8	3144.3
MHD Duct Exit	6	13.09	3460.4	3144.3
Diffuser Exit	7	17.00	3644.0	3144.3
Preheater Exit	8	16.52	2538.0	3435.8
Air Quench Chamber Exit	9	15.58	1880.0	3435.8
Steam Generator Exit	10	14.696	305.0	3435.8

Fig. 9.1–Schematic diagram and state points for open-cycle MHD Base Case ✓

heating value is assumed in the calculation. This heat is used in the steam-bottoming plant. Due to the high combustion temperature [2708°K (4415°F)], some of the slag will be vaporized and will not be removed through the slag tap. A carry-over of 20% has been assumed. Four combustor modules feed one mixer. The construction and cost of these are discussed in Appendix A 9.4. In order to prevent its being carried out with the slag, the potassium seed material will be injected in the mixer. Since the seed material is being used to remove sulfur, a level of 1% of the mass flow through the MHD generator was used in all cases to assure an adequate supply for the highest anticipated sulfur level. Furthermore, to provide this capability it is necessary to recycle some of the seed in a nonsulfate form. A treatment plant is provided to recover potassium from ash and to convert potassium sulfate to potassium carbonate. The basic equation of the conversion is

$$K_2SO_4 + CO_2 + 4H_2 \rightarrow K_2CO_3 + H_2S + 3H_2O. \qquad (9.1)$$

The hydrogen is produced by gasifying the coal in an oxygen-blown gasifier. When account is taken of the hydrogen produced by the shift reaction

$$CO + H_2O \rightleftarrows CO_2 + H_2, \qquad (9.2)$$

an energy input of 8.38 MJ/(kg mol potassium sulfate) is found to be the thermodynamic minimum for this reaction. There are other processes in the conversion which ideally produce net energy output (for example, the Claus plant, which converts the hydrogen sulfide to elemental sulfur), but these outputs are small. The irreversibilities of practical equipment are such that the actual energy requirement is about 30% larger than the minimum. The design, operation, cost, power, and energy requirements of this plant are discussed in Appendix A 9.1. Potassium to replace the stack losses and seed carried out in the slag is supplied in the carbonate form.

The pressurized [607.9 kPa (6 atm)] potassium-seeded combustion products leave the mixer through a converging nozzle where they are accelerated to a velocity of 775 m/s (2542 ft/s) before entering the active

MHD duct. This velocity (u) decreases in the direction of flow and is specified as a function of pressure and density as

$$\frac{d(u^2)}{dp} = 0.2\ \rho \tag{9.3}$$

with u in m/s, p in atm, and ρ in kg/m^3.

The magnetic field is imposed on the plasma by a superconducting magnet which is designed to limit cross-sectional nonuniformities to 5% and 8% in the directions parallel to and transverse to the field lines, respectively. The nominal magnetic field over the upstream portion of the MHD duct is 6 T. At the duct section where the pressure reaches 202.6 kPa (2 atm), the magnetic field begins tapering to reduce the Hall field to acceptable levels at the duct outlet end. The taper specified as

$$\frac{dB}{dp} = 2\ \text{T/atm} \tag{9.4}$$

The generator is a segmented design with the electrodes paired diagonally to eliminate Hall current effects (Appendix A 9.11). The generator-loading coefficient K (the ratio of load voltage to open-circuit voltage) is 0.82 at the upstream end of the duct and tapers in the direction of flow according to Equation 9.6. For an 1180 MW MHD output the dK/dp value would be 0.0205/atm

$$\frac{dK}{dp} = 0.0205/\text{atm} \tag{9.5}$$

The value of dK/dp is calculated from the equation

$$\frac{dK}{dp} = 0.05 - 0.0025\ \text{PMHD} \times 10^{-8} \tag{9.6}$$

with PMHD in watts. This equation was used for all points in this study.

The expansion through the duct is carried to a pressure level that will permit a recovery to 117.2 kPa (1.157 atm) with an 80% efficient diffuser. The diffuser walls are cooled by some of the combustion air.

Heat transfer to the duct-cooling water of 10% of the generated power is assumed, and viscous losses are included using a Fanning friction factor of 0.005. The cooling water used is the steam plant feedwater. This heat, transferred directly to the steam bottom cycle without any need for a coupling heat exchanger or intermediate fluid, is used by the steam plant.

On leaving the diffuser, the combustion products enter the main combustion air heater (heat recovery exchanger). The combustion air is heated to 1589°K (2400°F) while the combustion products are cooled to approximately 1650°K (2511°F). Due to the very high temperature of the exhaust products and the fact that they contain slag and seed, the heat recovery exchanger is a radiant design. In this design it is not necessary for the exhaust products to contact the tubes, so it is possible that the tube surfaces can be protected from fouling and/or corrosion by bathing them in recycled products or air. This design also leads to generous passage widths so that plugging is minimized and maintenance facilitated. Silicon carbide tubes are used for the highest temperature [above 1380°K (2040°F)] portion of this exchanger, and a high-nickel alloy for intermediate levels. The design and construction of this heat exchanger and diffuser are described in Appendix A 9.2.

The products next enter the bottoming plant (coupling) heat exchanger (steam generator). The first section exists in a temperature range where the potassium sulfate and ash will be condensing. This will be a radiant section with widely spaced tube platens. The tubes are covered with a ceramic. The surface temperature will be maintained above the freezing point of the materials to permit them to drain off and be tapped from the bottom of the cavity. When the gas stream temperature approaches the freezing point of the seed ash mixture [$\sim$ 1300°K (1881°F)], air will be injected into the stream to quench and freeze the seed. We have assumed that the same air which is required to complete combustion of the fuel (stoichiometric ratio of 1.05) will serve this purpose. If

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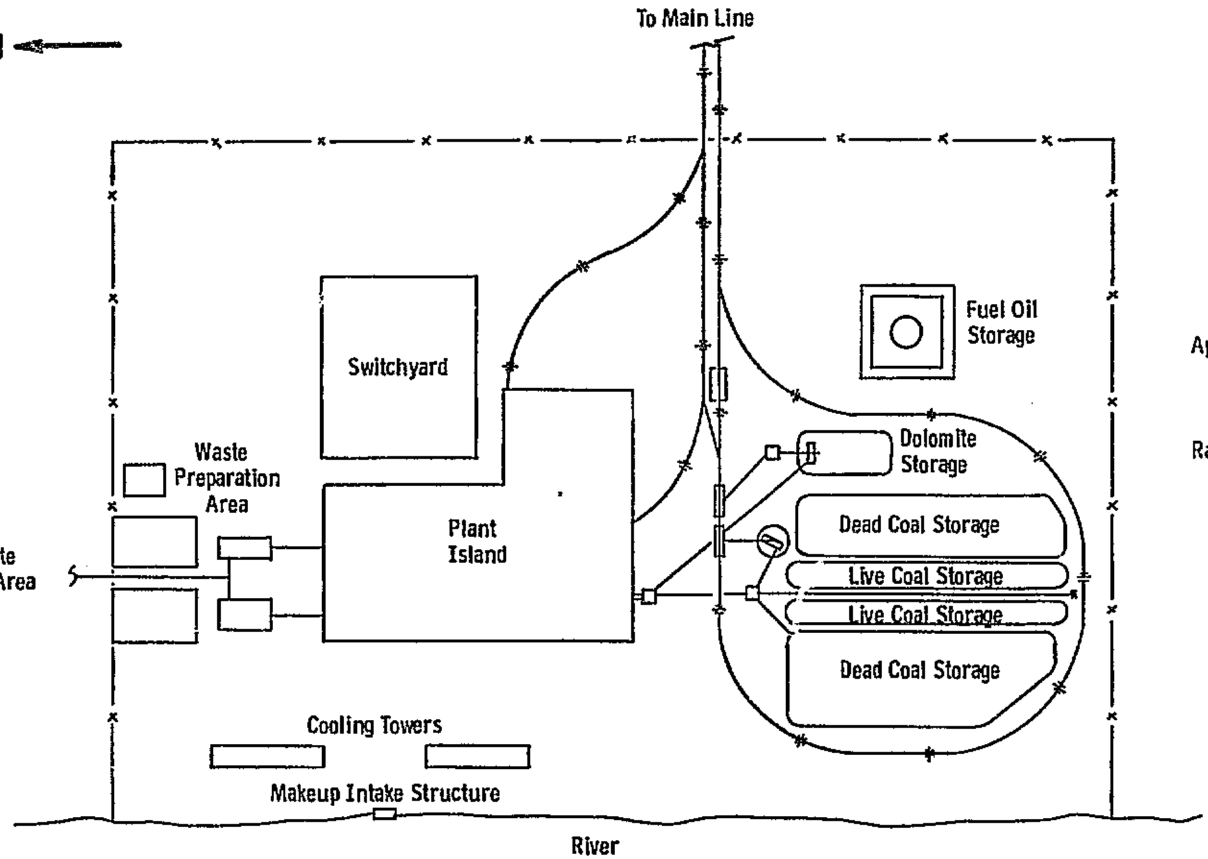

Approximate Site Area:
2600 Ft × 3880 Ft
~ 232 Acres

Railroad Requirements:
5 Miles to Main Line
1 Mile Passing Track
0.6 Mile Spurs
1.1 Miles of Loop Track
7.7 Miles Total Track

Fig. 9.2—Open-cycle MHD plant Base Case 2

Scale:
0 100 300 500

more quenching is required, some combustion products can be recirculated. After quenching, the seed and ash will be in an innocuous, dry, fluffy form. These gases will then be cooled to 425°K (306°F) by transferring heat to the reheat steam and low-temperature primary steam in convection tube banks. A detailed discussion of the steam generator design is included in Appendix A 9.3.

The products then pass to the electrostatic precipitator where the dry seed and ash products are removed from the gases. The selection and design of precipitators for this job are discussed in Appendix A 9.12. The required precipitator efficiency was determined from the current particulate emission limitation imposed by the EPA. These limits are more stringent than the commonly assumed economic seed recovery values. The calculations of required efficiency are described in Appendix A 9.1. The exhaust gases are then exhausted to the stack of the plant.

The precipitator dust and the seed tapped from the bottom of the superheater pass through a seed treatment plant. Here, sufficient potassium to react with the sulfur in the coal stream is converted from the sulfate to the carbonate form. The treatment plant produces high-quality elemental sulfur. The design and energy requirement calculations for this treatment plant are discussed in Appendix A 9.1.

The bottoming plant for this case is a 24.2 MPa/811°K/811°K (3500 psig/1000°F/1000°F) steam plant with a condenser pressure of 6.75 kPa (2 in Hg) abs. This back pressure was chosen as being reasonably compatible with ISO ambient conditions and evaporative cooling towers. One steam turbine drives the compressor to supply the air to the MHD plant. The remaining steam is used in a conventional turbine-generator to provide ac power.

Figure 9.2 shows the assumed 232-acre site, including scaled size blocks representing the plant island area, coal and oil storage and handling facility, switchyard, cooling towers (not to scale), and on-site railroads. The waste storage area is not shown. Some details of the balance of plant assumed may be found in the amount column of the detailed accounts listing (Table 9.10). A plant island layout and elevation is included in Appendix A 9.6 as Figure A 9.6.2.

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TABLE 9.1 – OPEN-CYCLE MHD PARAMETRIC INVESTIGATION OF BASE CASE 2

Parametric Point	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17
Power Output, MWe	1993	1191	581	1989	1911	1969	1952	1970	1993	1990	1993	1981	1967	1978	1966	1966	1988
Fuel Type																	
Bituminous Coal, As Rec'd.	X	X	X						X		X	X	X	X	X	X	X
Bituminous Coal, Min. Dry ①				X						X							X
Subbituminous Coal, Min. Dry					X												
Subbituminous Coal, Max. Dry ①						X											
Lignite Coal, Min. Dry							X										
Lignite Coal, Max. Dry								X									
Combustor Stages	1	1	1	1	1	1	1	1	2	2	3	1	1	1	1	1	1
Ash Carryover, $	20	20	20	20	20	20	20	20	10	10	5	100	20	20	20	20	20
Diluent Exhaust Gas Used													X				
Preheat Temp., °F	2398	2398	2398	2402	2404	2404	2402	2384	2398	2402	2398	2398	2402	2400	2390	2398	2400
Combustor Temp., °F	4414	4414	4414	4486	4387	4351	4324	4220	4414	4486	4414	4414	4400	4443	4463	4414	4503
Combustor Press., psia	88.2	88.2	88.2	88.2	88.2	88.2	88.2	88.2	88.2	88.2	88.2	88.2	88.2	117.6	88.2	88.2	102.9
Steam Throttle Press., psia	3500	3500	3500	3500	3500	3500	3500	3500	3500	3500	3500	3500	3500	3500	3500	2400	3500

Note:

① Min. Dry and Max. Dry Refer to Minimum Drying and Maximum Practicable Drying Respectively

Base Case 2 and its variations are summarized in Table 9.1. The nominal power output of the base case is 2000 MWe.

9.2.2 Variations on Base Case 2

Points 2 and 3 have nominal power outputs of 1200 and 600 MWe, respectively. All other points have a nominal output of 2000 MWe.

For Point 4, the fuel is the bituminous coal with minimum drying (moisture level reduced from 13 to 3%). The fuels for Points 5, 6, 7, and 8 are the subbituminous coal with 20% moisture, the subbituminous coal with 16% moisture, lignite with 27% moisture, and lignite with 18% moisture, respectively.

Points 9 and 10 use two-stage combustors with the bituminous coal, with 13% and 3% moisture, respectively. These combustors are discussed more fully under Base Case 1 and in Appendix A 9.4. The slag carry-over is assumed to be 10% for these two cases.

Point 11 has a three-stage combustor burning the bituminous coal as received. The slag carry-over is assumed to be 5% with this combustor. Point 12 has a one-stage combustor with no slag removal so that all of the ash in the coal is carried over into the plant.

Point 13 differs from Base Case 2 in that here enough exhaust products are recycled with the combustion air to limit the combustor temperature to 2700°K (4400°F). This concept is discussed more fully under Base Case 1. Due to the moisture level of the coal and the relatively low air-preheat temperature, only 1.3% of the gas flow through the MHD duct is recycled.

Points 14 and 15 have duct pressure ratios of 8 and 10, respectively rather than the base value of 6.

Point 16 has a 16.6 MPa (2400 psig) steam plant in place of the 24.2 MPa (3500 psig) of the base case.

Point 17 combines the drier bituminous coal with a duct pressure ratio of 7.

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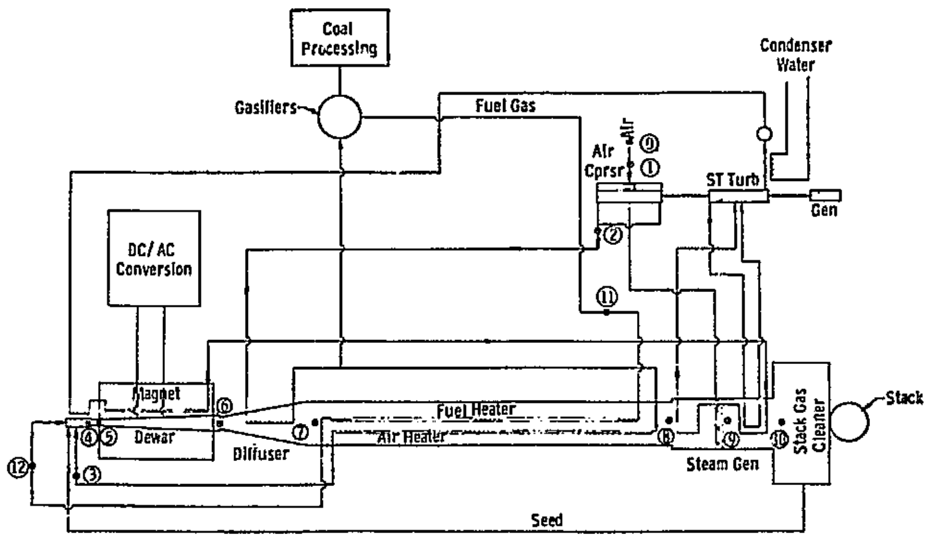

Location	Point No.	Pressure, Psia	Temperature, °F	Flow, lb/s
Ambient	0	14.696	59.0	1710.2
Compressor Inlet	1	14.40	59.0	1710.2
Compressor Outlet	2	158.94	603.8	1710.2
Air Preheater Outlet	3	154.31	2587.4	1710.2
Combustor Outlet	4	146.96	4400.0	3030.2
MHD Duct Entrance	5	99.02	[illegible]	3030.2
MHD Duct Exit	6	13.54	[illegible]	3030.2
Diffuser Exit	7	17.00	3446.0	3030.2
Air Preheater Exit	8	16.52	2230.4	3210.2
Air Quench Chamber Exit	9	15.58	1880.0	3210.2
Steam Generator Exit	10	14.696	305.0	3210.2
Fuel Preheater Entrance	11	158.90	1600.0	1282.9
Fuel Preheater Exit	12	154.31	2591.0	1282.9

Fig. 9.3—Schematic diagram and state points for open-cycle MHD Base Case 3

9.2.3 Base Case 3

Base Case 3 includes the use of a low-Btu gas produced from the bituminous coal using the Westinghouse Fluidized Bed Gasifier and high-temperature desulfurization. The operating conditions of the gasifier and the gas properties are given in Table 2.8. The gasifier is integrated into the MHD cycle, as indicated in the schematic diagram (Figure 9.3).

Some of the air delivered by the main air compressor is preheated to 672°K (750°F) and taken to the gasifiers to react with the raw coal. The clean product fuel gas is then preheated by heat recovered from the MHD exhaust products. It is then burned with the remaining combustion air. The thermodynamic calculations were made assuming equal preheat temperatures for the air and fuel gas. Due to materials problems (Appendix A 9.2), differing temperatures were used in the heat exchanger design.

The seed material [cesium carbonate (Cs_2O_3)] is injected into the combustor at a rate equal to 1% of the gas flow rate through the duct. Makeup cesium is supplied as an ore containing 25% cesium.

The duct expansion parameters are specified or calculated as discussed under Base Case Number 2.

On leaving the diffuser the MHD exhaust gases pass through the fuel gas and air preheaters. The design and cost of these preheaters are discussed in detail in Appendix A 9.2.

The heat recovery steam generator design is discussed in Appendix A 9.3. The temperature at which the quench air should be injected will be influenced by the fact that the material to be quenched will be nearly pure cesium carbonate. When the gases have been cooled to 425°K (306°F), the cesium carbonate will be recovered from the stream by an electrostatic precipitator whose efficiency is again determined by current particulate emission limits. Since the seed will be reasonably clean cesium carbonate, it will be recycled with a minimum of treatment (possibly grinding or pulverizing).

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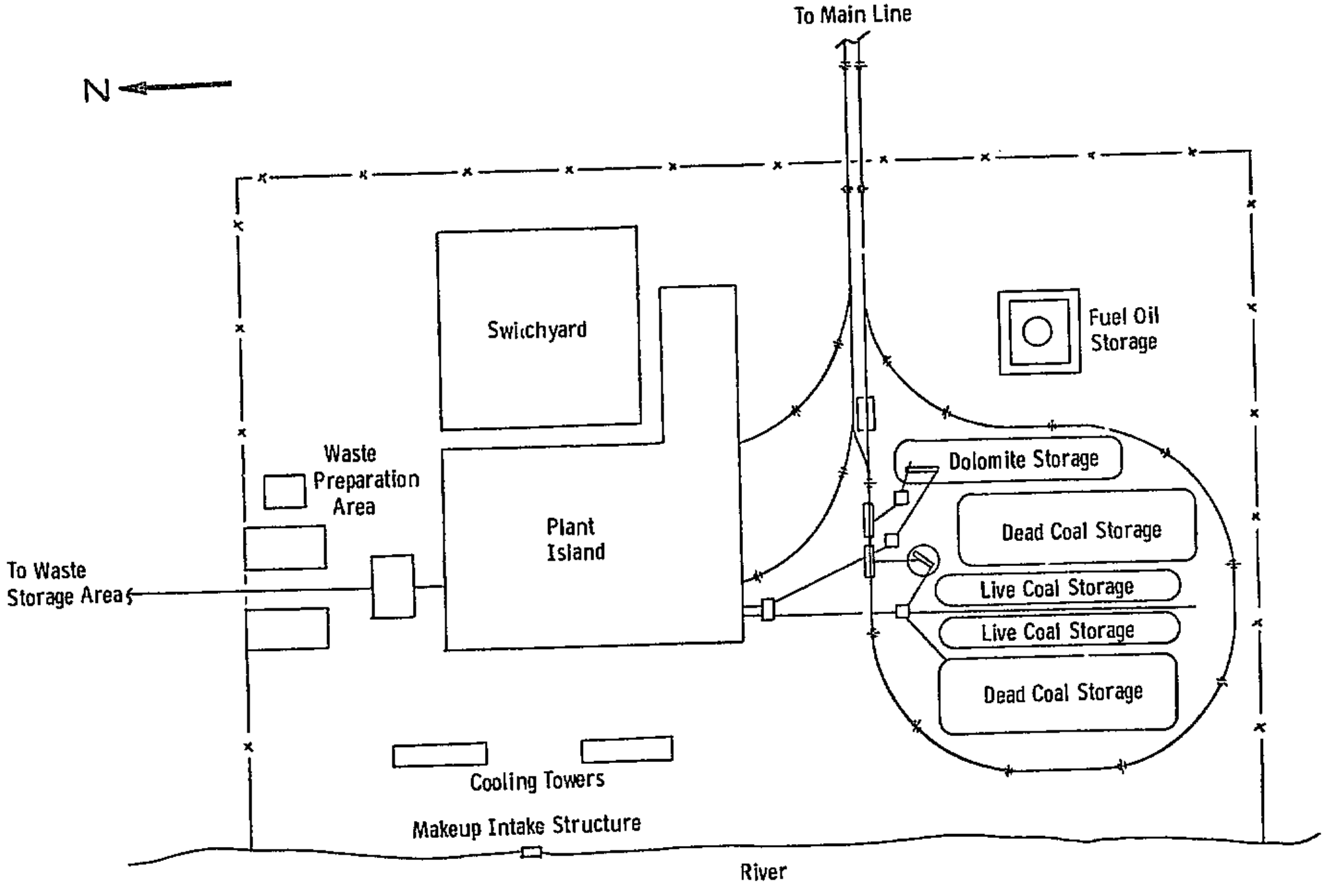


Approximate Site Area:
2600 Ft × 3920 Ft
~ 234 Acres

Railroad Requirements:
5 Miles to Main Line
1 Mile Passing Track
0.6 Mile Spurs
1.1 Miles of Loop Track
7.7 Miles Total Track

Fig. 9.4—Open-cycle MHD plant Base Case 3

Scale:
0 100 300 500

The steam-bottoming plant for this case has the same parameters as Base Case 2. However, due to the higher pressure ratio (10:1), a larger portion of the steam plant output is required to drive the air compressor. The nominal output of this case is 2000 MWe.

The site layout for Base Case 3 is similar to that for Base Case 2 and is shown in Figure 9.4. The accounts listing is given as Table 9.16. The plant island layout and elevation are included in Appendix 9.6 as Figure A9.6.3.

9.2.4 Variations on Base Case 3

Base Case 3 and its variations are summarized in Table 9.2.

Points 2 and 3 have nominal plant outputs of 1200 MWe and 600 MWe respectively.

In Point 4, the air and fuel preheat temperature are raised to 2028°K (3191°F). The corresponding flame temperature at 1.013 MPa (10 atm) is 2855°K (4680°F).

In Point 5, the duct pressure ratio is raised to 15 with essentially the same air and fuel preheat temperature [2025°K (3186°F)] as Point 4.

9.2.5 Base Case 1

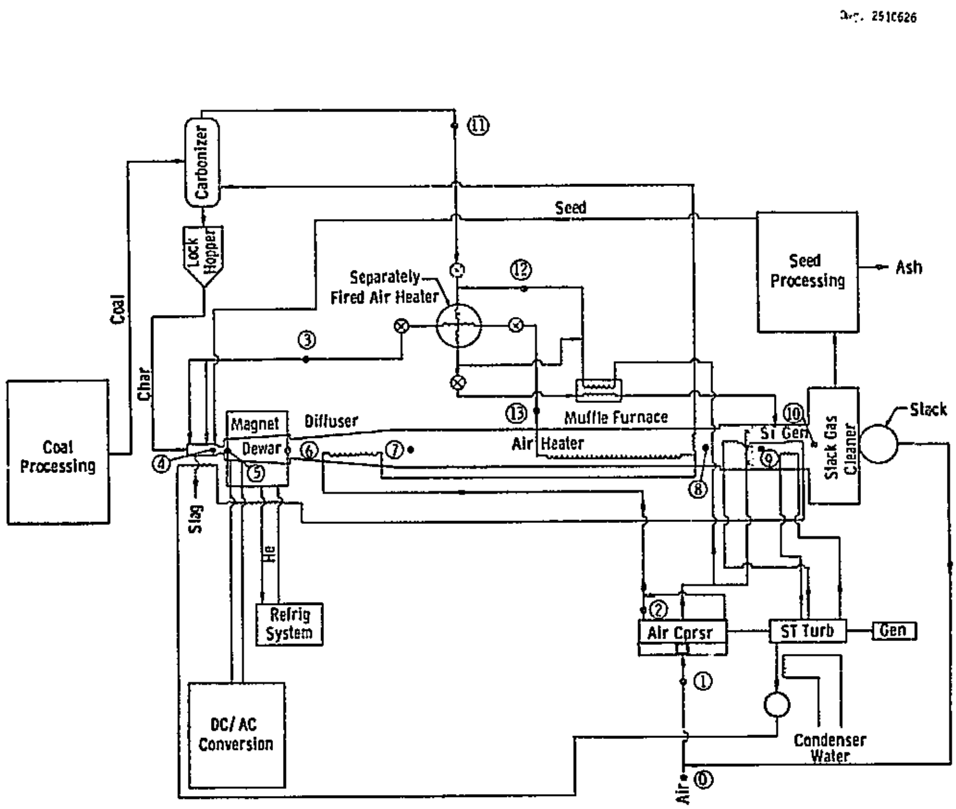
In Base Case 1 , the coal is carbonized to produce a char (which is a very desirable fuel for the MHD duct) and a clean low-Btu fuel gas (gapor). As shown in the schematic (Figure 9.5), the char is introduced into a two-stage slagging MHD combustor while the gapor is burned in a separate stove to preheat a mixture of combustion air and recycled exhaust products.

The first stage of the MHD combustor is operated with considerably less than stoichiometric oxygen to limit the temperature levels so that most of the slag will be liquid and can be tapped off. In the second stage the remaining mixture of air and recycled exhaust products and the seed are introduced. The second stage is operated at 95% of stoichiometric oxygen to limit production of nitric oxides. The mixture

Dwg. 1675B40

TABLE 9. 2 — OPEN-CYCLE MHD PARAMETRIC INVESTIGATION OF BASE CASE 3

Parametric Point	1	2	3	4	5
Power Output, MWe	1885	1131	566	1889	1900
Fuel					
Bituminous Coal, Gasified	X	X	X	X	X
Preheat Temperature, °F	2587	2587	2587	3190	3180
Combustor Temp., °F	4400	4400	4400	4679	4724
Combustor Press., psia	147.0	147.00	147.0	147.0	220.4
Steam Throttle Press., psig	3500	3500	3500	3500	3500



Location	Point No.	Pressure, Psia	Temperature, °F	Flow, lb/s
Ambient	0	14.696	59.0	1976.3
Compressor Inlet	1	14.40	132.1	2934.7
Compressor Outlet	2	95.36	54.69	2934.7
Air Preheater Outlet	3	92.58	2933.5	2934.7
Combustor Outlet	4	88.18	4400.0	3194.8
MHD Duct Entrance	5	58.59	4162.4	3194.8
MHD Duct Exit	6	12.94	3422.6	3194.8
Diffuser Exit	7	17.00	3620.6	3194.8
Air Preheater Exit	8	16.52	2483.0	3194.8
Air Quench Chamber Exit	9	15.58	1880.0	3493.4
Steam Generator Exit	10	14.696	305.0	3493.4
Gapor	11	16.16	800.0	113.0
SFA Combustion Air	12	21.25	2384.0	632.0
Cross-over Air	13	93.5	2384.0	2934.7

Fig. 9.5—Schematic diagram and state points for open-cycle MHD Base Case 1

of air and recycled products is adjusted to obtain flame temperatures of 2700°K (4400°F) at a combustor pressure of 607.8 kPa (6 atm) with an air and recycled products preheat temperature of 1885°K (2933°F), a heat loss to the slag and combustor walls of 5% of the higher heating value of the fuel, and sufficient seed (as potassium carbonate and potassium sulfate) to produce a 1% potassium level in the MHD duct. The design and operation of this combustor are discussed in detail in Appendix A 9.4.

The parameters for the expansion through the MHD generator are all specified as described for Base Case 2, and (with two exceptions) the MHD exhaust products are handled the same as in Base Case 2.

The first exception is that the products of combustion of the gapor are mixed with the MHD exhaust products. These vapor products can be used as the protective layer around the heat recovery exchanger tubes and/or to supplement the air used in quenching the seed and slag materials.

The other exception is that some of the products are directed back from the stack to the inlet of the main air compressor. For the base case (Point 1) the recycled product flow rate is 30% of the gas flow through the MHD generator, or 435 kg/s (979 lb/s).

After passing through the compressor, the mixture is heated to 1580°K (2385°F) by heat recovered from the MHD exhaust stream as in Base Case 2. The mixture is then heated further to 1885°K (2934°F) in the stoves (periodic-type heat exchangers). These stoves are similar to those used in the steel industry, and their design and operation is described in detail in Appendix A 9.2. The use of the clean gapor as a fuel makes it possible to utilize this type of stove without plugging the passages in the brickwork unduly.

In order to make the energy of the gapor available at a usable temperature level [> 1580°K (2385°F)] it is necessary to preheat its combustion air. The air is preheated regeneratively to 1580°K (2385°F) by the gapor products. This is done in a ceramic muffle furnace which is also described in Appendix A 9.2. Due to its agglomerating tendency, it is necessary to preoxidize the bituminous coal before introducing it to

the carbonizers. During this step, all of the moisture is removed from the coal. As a result, the gapor produced has a relatively high heating value 13.532 MJ/kg (5819 Btu/lb). If this fuel is burned with 1580°K (2385°F) air, the resultant flame temperature will create severe materials problems in the hot end of the stove. To limit the flame temperature to 2255°K (3600°F), 1.87 kg of products are recycled from the exhaust of the stove for each kilogram of combustion air. To accomplish this the pressure of the recycled products must be increased by an amount greater than the pressure drop through the stove (by jet pumping with the combustion air). The inherent problems in this procedure is and some alternative methods are discussed in Appendix A 9.5. The products, which are exhausted from the stoves, heat the combustion air in the muffler furnace and are then injected into the MHD exhaust stream.

The base case requires 40 of these stoves, and valves are required on the incoming and outgoing streams to permit cycling of the stoves. To minimize the number of valves, a complex manifolding scheme was developed and is discussed in Appendix A 9.6.

The nominal power output of Base Case 1 is 2000 MWe. The site plan for Base Case 1 is shown in Figure 9.6 and includes a dolomite-handling facility. The accounts listing for Base Case 1 is presented as Table 9.22. The plant island layout and elevation are included as part of Appendix A 9.6 as Figure A 9.6.1.

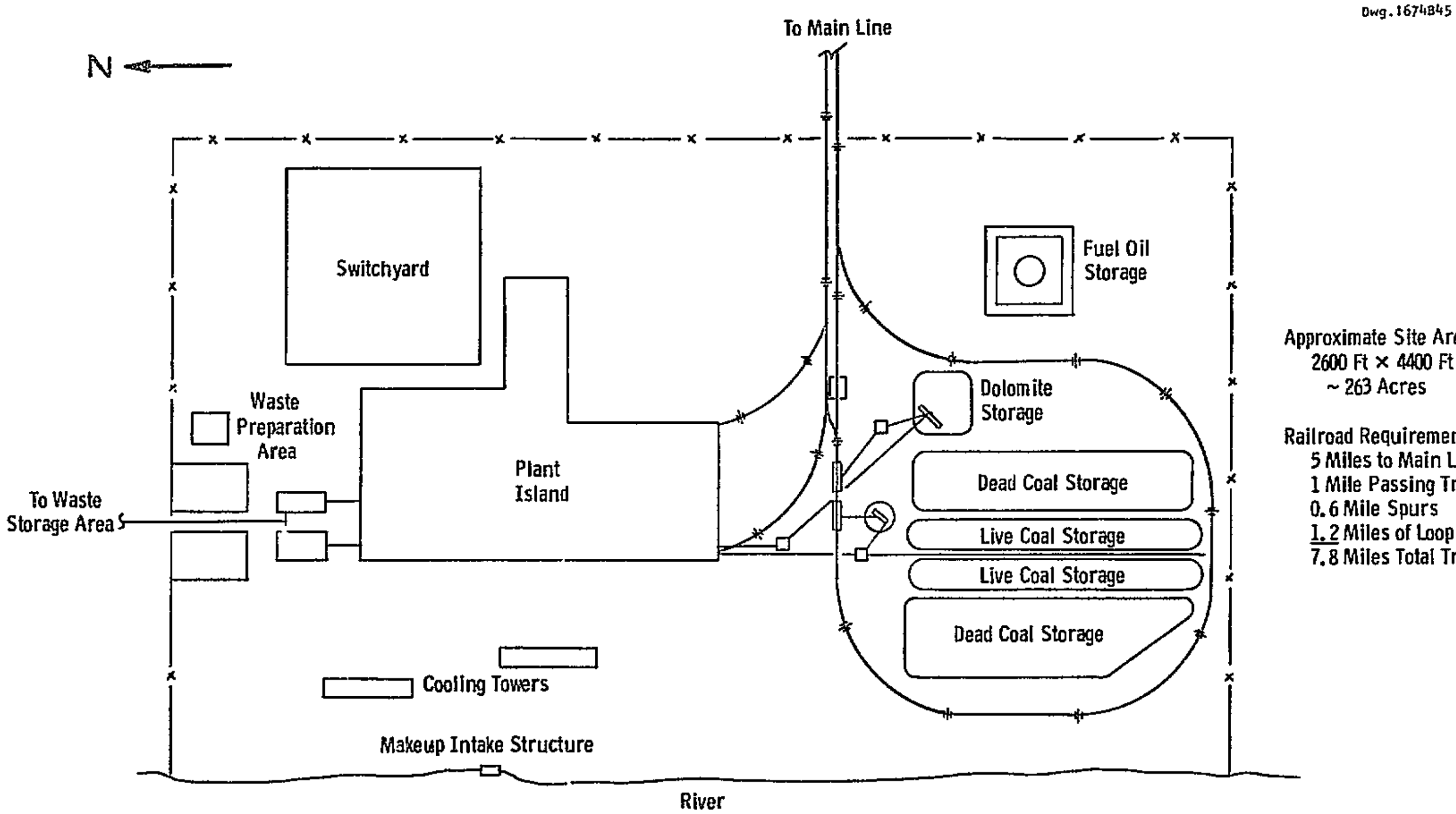
9.2.6 Variation on Base Case 1

Variations on Base Case 1 are summarized in Table 9.3.

Points 2 and 3 have nominal outputs of 1200 and 600 MWe respectively.

Point 4 was to have used the bituminous coal as received. The need for preoxidation, however, eliminated the distinction between this case and the base case.

Points 5, 6, 7, and 8 were to have been similar to the base case with the subbituminous coal with 20% moisture, the subbituminous coal with 16% moisture, the lignite with 27% moisture, and the lignite with 18%



Approximate Site Area:
2600 Ft × 4400 Ft
~ 263 Acres

Railroad Requirements:
5 Miles to Main Line
1 Mile Passing Track
0.6 Mile Spurs
1.2 Miles of Loop Track
7.8 Miles Total Track

Fig. 9.6—Open-cycle MHD plant Base Case 1

Scale:
0 100 300 500

Dwg 1675041

TABLE 9.3. – OPEN-CYCLE MHD PARAMETRIC INVESTIGATION OF BASE CASE 1

Parametric Point	1	2	3	5	6	7	8	9	10	11	12	13	14	15	16	17
Power Output, MWe	1971	1172	593	1930	1939	1932	1923	1970	1971	1961	1955	1917	1981	1979	1971	1943
Fuel																
Bituminous Coal, Preoxidized	X	X	X					X	X	X	X	X	X	X	X	X
Subbituminous Coal, Min. Dry (1)				X												
Subbituminous Coal, Max. Dry (1)					X											
Lignite Coal, Min. Dry						X										
Lignite Coal, Max. Dry							X									
Combustor Stages	2	2	2	2	2	2	2	3	1	1	2	2	2	2	2	2
Ash Carryover, %	10	10	10	10	10	10	10	5	20	100	10	10	10	10	10	10
Diluent Exhaust Gas Used	X	X	X	X	X	X	X	X	X	X		X	X	X	X	X
Preheat Temp., °F	2993	2993	2993	2614	2591	2388	2357	2993	2993	2993	2931	2542	3532	2933	2933	2993
Combustor Temp., °F	4400	4400	4400	4400	4400	4400	4356.8	4400	4400	4400	4855.4	4400	4580	4400	4400	4400
Combustor Press., psia	88.2	88.2	88.2	88.2	88.2	88.2	88.2	88.2	88.2	88.2	176.4	88.2	117.6	117.6	147.0	88.2
Steam Trhottle Press., psia	3500	3500	3500	3500	3500	3500	3500	3500	3500	3500	3500	3500	3500	3500	3500	2400

Note:
(1) Min. Dry and Max. Dry Refer to Minimum Drying and Maximum Practicable Drying Respectively

moisture respectively as fuels. Since preoxidation of these coals was not necessary, however, their gapors contained large quantities of moisture and they were inferior fuels. As a result, it was found that if the gapor combustion air and the stream to be preheated entered the stoves at 1580°K (2385°F), the energy available above the temperature of the steam to be preheated [plus a 100°K (180°F) differential to provide for heat transfer] resulted in temperature rises of the preheated stream of less than 40°K (72°F). This is illustrated in Point 7 where, when both streams entered the stoves at 1525°K (2286°F), a rise of 40°K (72°F) was achieved with no recycled products. To avoid this situation, we reduced the temperature of both streams entering the stoves for Points 5, 6, and 8 to 1350°K (1970°F). At this level it may be possible to eliminate the need for ceramic heat exchangers in the MHD exhaust stream. The 1525°K (2286°F) level was chosen for Point 7 to determine its overall effect on cycle performance and cost. To obtain a 40°K (72°F) rise, it was necessary to eliminate the recycle stream for Point 7. The resultant final temperatures of the preheated streams are 1695, 1708, 1580, and 1565°K (2592, 2615, 2358, and 2385°F), and the recycled products are 14, 14, 0, and 7 percent for Points 5, 6, 7, and 8 respectively. The MHD combustor flame temperature for Point 7 is 2676°K (4357°F). There is no recycling of products around the stoves, since flame temperatures are within chrome-alumina brick capabilities.

Points 9, 10, and 11 differ from the base case in that a three-stage combustor is used for Point 9 and a single-stage combustor for Points 10 and 11. Ash carry-overs of 5, 20, and 100% are assumed for Points 9, 10, and 11, respectively.

In Point 12 no exhaust gas is recycled. The final preheat temperature is held at the same level as the base case by reducing the temperatures of the air to be preheated by the stoves and the stove combustion air preheat temperature to 1438 and 1050°K (2129 and 1430°F) respectively. The MHD combustor temperature is 2953°K (4856°F), and a pressure level of 1.2159 MPa (12 atm) is used to match this temperature level.

In Point 13, the final preheat temperature is reduced to 1668°K (2543°F) with corresponding reductions in the temperatures of the combustion air and preheat stream entering the stoves to [630 and 1328°K (675 and 1931°F)] respectively. For this case the MHD generator stream contains 18% recycled products to limit the combustor temperature to 2700°K (4400°F). To limit the stove flame temperature to 2255°K (3600°F), 0.7 kg of gapor combustion products are recycled for each kilogram of combustion air.

In Point 14, a final preheat temperature of 2218°K (3533°F) is used. To reach this temperature, the preheat stream and gapor combustion air enter the stove at 1970°K (3087°F). To reach the final preheat temperature, it is necessary to limit the recycled products flow and allow the MHD combustor temperature to increase to 2800°K (4581°F). At this level, 41% of the MHD duct gas flow is recycled products. It is also necessary to remove the limitation on the combustion temperature of the gapor. No gapor products are recirculated, and the gapor combustion temperature is 2505°K (4050°F). In calculating the cost of the stoves, it was assumed that zirconia brick would be used where the gas temperature exceeded 2255°K (3600°F).

Points 15 and 16 have MHD pressure ratios 8 and 10 respectively. Because of reduced dissociation, it is necessary to recycle 32 and 33% (as a fraction of flow in the MHD duct) of the products to maintain the 2700°K (4400°F) MHD combustor.

Point 17 has a 16.6 MPa/811°K/811°K (2400 psig/1000°F/1000°F) steam bottoming plant.

9.3 Approach

The basis of the open-cycle MHD duct calculations is a Westinghouse proprietary computer program (MHD-2502) which calculates the equilibrium composition, the thermochemical properties, and the electrical conductivity of seeded products of combustion. This program requires that the fuel and oxidant components be given as chemical compounds, that their composition be given as mole fractions of the compounds, and that their energy content be given as heats of formation of the compounds. A simple auxiliary program (INPUTAFLOWS) was written to convert coal and

oxidant compositions (which were given as percents by weight) into the desired form and punch cards which are used as the input for MHD-2502. This program listing is in Appendix A 9.7.

The MHD 2502 program was modified so that the results of its calculations were printed on a computer file as well as on paper. Another auxiliary program (MAP DISK) was written to take selected data (specific enthalpy, electrical conductivity, specific entropy, etc.) from this file and organize them into rectangular arrays of fluid properties. The array arguments are temperature (°K) and pressure (atm). A listing of this program is in Appendix A 9.7.

The arrays then serve as data for various versions of the MHD-DUCT program. These duct programs calculate the performance and design of the MHD duct and estimate the performance of a combined MHD-bottoming plant. The MHDDUCT program was originally written for another contract, but extensive modifications were required for this work and three different versions were used for this contract.

9.3.1 Duct Program for Base Case 2

The core of the duct programs is a finite element solution of the MHD generator performance. Using the data generated by the MHD 2502, the duct programs calculate the element of duct length required for a specified small pressure step. For the small step the properties of the fluid are assumed to be constant and uniform over the cross-section.

The duct calculation begins with a specified combustion pressure (PCOMB)* and a specified combustion temperature (TCOMB)and calculates the static pressure and temperature at the start of the generator for the inlet velocity (UO). It is terminated when a pressure level is reached at which a diffusion of specified efficiency will result in a specified duct outlet pressure (diffuser exit pressure). The calculation is initially performed for an estimated flow of gas, and an iteration is carried out to determine the gas flow rate required for a specified MHD power output (PE).

*The nomenclature used in the duct program is defined in Table 9.4.

Table 9.4 - Nomenclature Used in Open-Cycle MHD Duct Program Outputs

A4	= Area of duct cross-section at upstream end, m^2
A5	= Area of duct cross-section at downstream end, m^2
B	= Magnetic field intensity on duct centerline, Wb/m^2
C	= Velocity coefficient $[d(U^2)/dp = 2 \quad C/Density]$
D	= Height-width of generator, m
DB	= Magnetic field decrement ($dB/dp = DB$, Wb/m^2-atm) applied only when pressure is below 2 atm
DK	= Generator coefficient decrement ($dK/dp = DK$, atm^{-1}) presently calculated as $DK = .05 - .0025 \quad PMHD \times 10^{-8}$
DM	= Mean height-width of generator $(D4 + D5)/2$, m
D4	= Height-width of generator at inlet end $(\sqrt{A4})$, m
D5	= Height-width of generator at outlet end $(\sqrt{A5})$, m
ETAMHD	= Conversion efficiency of MHD generator $(\Delta H/\Delta H_s)$
HR	= Plant heat rate, Btu/kWh
K	= Generator segment loading coefficient operating voltage/open circuit voltage)
M	= Mass flow rate of gas, kg/s
MA	= Mass flow rate of dry air through duct, kg/s
MAW	= Mass flow rate of moist air through duct, kg/s
MA'	= Mass flow rate of supplementary air, kg/s
MC	= Mass flow rate of dry ash free combustible, kg/s
MCW	= Mass flow rate of moist combustible, kg/s
MG	= Mass flow rate of gas through MHD duct, kg/s
MS	= Mass flow rate of seed compound, kg/s

Table 9.4 (continued)

MU	= Electron mobility, m^2/Wb
MWC	= Mass flow rate of moisture in coal, kg/s
MWS	= Mass flow rate of water with seed, kg/s
N	= Conversion efficiency of generator segment
NA	= Ratio of moles of air to moles of oxidizer
NC	= Ratio of moles of combustible to moles of fuel
P	= Pressure in gas states or generator design (atm) or net plant power output, W
PAUX	= Auxiliary power requirement, W
PC	= Main air compressor power, W
PCOMB	= Combustor pressure, atm
PE	= Electrical output of MHD generator, W
PHI	= Equivalence ratio—flow rate of air divided by stoichiometric flow rate of air
PMHD	= Net power output of inverters, W
PS	= The electrical output of a generator receiving all the mechanical power of the steam turbine, W
PSE	= The electrical output of a generator receiving the output of a turbine which also drives the air compressor, W
PT	= The mechanical power available for the generator after deducting the power for the air compressor, W
P1	= Compressor outlet pressure, atm
QA	= Heat added to the oxidant (air + recycled products) in the air preheater, J
QS	= Total heat transferred to the steam cycle, W

Table 9.4 (continued)

QSO	=	Heat transferred from the exhaust gases to the steam, W
QS1	=	Heat transferred from the combustor to the coolant, W
QS2	=	Heat transferred from the MHD duct to the coolant, W
RHO	=	Gas density, kg/m^3
RMAP	=	Ratio of supplementary air flow rate to MG
RMC	=	Ratio of dry coal flow rate to MG
RMCW	=	Ratio of moist coal flow rate to MG
RMOX	=	Ratio of oxygen flow rate to MG
RMRP	=	Ratio of recirculated gas flow rate to MG
RMS	=	Ratio of seed compound flow rate to MG
RMWC	=	Ratio of coal moisture flow rate to MG
RMWS	=	Ratio of seed water flow rate to MG
S	=	Gas entropy, J/kg-°K
SIGMA	=	Gas conductivity, mhos/m
T	=	Temperature, °K
TCOMB	=	Combustor temperature, °K
TSTACK	=	Temperature of gases at exit of bottoming plant heat exchanger, °K
THETA	=	Heating value of dry ash free fuel, J/kg
U	=	Gas velocity, m/s
UO	=	Gas velocity at inlet end of duct, m/s
XS	=	Stoichiometric ratio of oxidant moles to fuel moles

The MHDDUCT programs also calculate:

- The pressures at all stations defined in the cycle schematics. These are calculated from a specified ambient pressure (station 0), combustor pressure, and component pressure drops.
- The mass flow rates of fuel, air, moisture and seed through the MHD duct from the calculated gas flow rate and the flow ratios RMAP, RMC, RMCW, RMOX, RMRP, RMS, RMWC, and RMWS. These ratios are calculated by INPUTAFLOWS (Appendix A 9.7) for the first two versions of the duct program (FRECIRCINJOX and DQHDUCT). For CHRDUC2, they are calculated in a subroutine of the program (PRELIM).
- Compressor power (PC) required for the moist air (MAW), recycled products flow rate, and the compressor pressure ratio.
- The air preheat temperature and the heat required for it (QA). QA is the heat which must be added to the reactants if they are to reach a given flame temperature with the specified combustor heat loss. It is based on the output of MHD 2502 and is calculated in the duct programs.
- The heat which is available to the bottoming plant (QS). This is calculated by deducting QA from the total heat available by cooling the MHD exhaust products to TSTACK plus the combustor, duct generator, and inverter heat losses. In the program this was achieved by using the MHD 2502 program to calculate the composition and molecular weight of the products after injection of the supplementary air (MA'). Dr. S. W. Way, however, has pointed out that these results do not include the heat released during conversion of potassium carbonate to potassium sulfate

and subsequent condensation and solidification of the potassium sulfate. To correct for this omission Dr. Way calculated the correct enthalpy of the exhaust products ducts by hand, and his results have been used to correct the value of QS obtained from the program.

- The duct program then calculates the net plant output (P) for a range of bottoming plant efficiencies by adding the bottoming plant power to the net MHD power (PMHD) and subtracting the compressor power (PC).
- The thermodynamic efficiency of the plant (overall efficiency) or heat rate (HR) is then estimated by calculating the heat input from the product of the combustible flow rate (MC) and the heating value (THETA) divided by the plant output (P).

A listing of the duct program used for Base Case 2 and its variations (PRECIRCINJOX) is included in Appendix A 9.7. Table 9.5 is the duct program output for Base Case 2.

9.3.2 Duct Program for Base Case 3

Since Base Case 3 uses an integrated low-Btu gasifier, it was necessary to modify the program somewhat. The additional output quantities are listed and defined in Table 9.6. Table 9.7 is the output for Base Case 3. The listing of this program (DQHDUCT) is in Appendix A 9.7.

The changes in the duct program were required to account for the fact that:

- Some of the compressed air is needed to gasify the fuel and, hence, appears as part of the dry ash free combustible flow. The airflow rates include only the air to the MHD combustor.
- Some of the exhaust heat is needed to generate low-pressure steam for the gasifier.

Table 9.5   Output of FRECIRCINJOX for Base Case 2

```
PATCHES  2/5/75  FOR PWRE FUNCT3                                                              DATE 022575          PAGE

     PROD. OF ILL. #6 COAL AS REC'D. WITH 1% POT. SEED AS DRY CARBONATE
     OXIDANT IS AIR WITH .639% MOIST. (59F-60% REL. HUM.)

PHI =     .95     PMHD =  1.18000+09     TCOMB =  2708.0     PCOMB =    6.000     UO =  775.00
RMRP =        .000       RMOX =          .000

RMAW =    8.804800-01       RMC =    8.720270-02       RMS =    1.767260-02
RMWC =    1.464650-02       RMMS =    .000000         RMCN =    1.018500-01       RMAP =    9.268190-02

WC =    15.60      NC =    .8712       NA =    .9861      HHVDAF =    1.393800+04
XS =    5.023440+00       T(STACK) =    4.250000+02

EFFICIENCIES
    DIFFUSER                        .8000
    ROTATING GENERATOR              .9840
    DC/AC INVERTER                  .9850

HEAT TRANSFER RATIO
    FROM COMBUSTER TO SUBPOSED PLANT    =      .050
    MHD GENERATOR TO SUBPOSED PLANT     =      .100

    FRICTION FACTOR IN MHD DUCT =      .0050
    PAUX/PSE =      .0150
    (PI-PCOMB)/PCOMB =      .0E15
    PC/MAW =    2.29918+05
    QA/MAW =    1.23412+06
    THETA =        3.24195+07
    PMHD =         1.18000+09
    PE =          1.19797+09 <-

   C =      .1000
   DK =      .0205
   DB =     2.0000
   HG    1.418334+03
   HG    1.426603+03

AIR SIDE            TOTAL    PRESSURE     TEMPERATURE

  COMPRESSOR INLET           .9800        288.3003
  COMPRESSOR OUTLET         6.4690        513.6719
  AIR PREHEATER EXIT        6.3000       1588.2133

GAS SIDE                      TOTAL    PRESSURE     TEMPERATURE

  MHD DUCT INLET                       6.0000       2708.0000
  DIFFUSER EXIT                        1.1573       2280.4177
  INJECTOR EXIT                        1.1236       2237.6417
  AIR PREHEATER EXIT                   1.0600       1625.8482
  BOTTOMING HEAT EXCH. EXIT            1.0000        425.0000
```

GAS STATES

POINT,	P	T	H	RHO	S	M
0	1.000	288.	7.			1256.134
1	.980	288.	7.			1256.134
2	6.469	514.	105.			1256.134
3	6.300	1588.	636.			1256.134
4	6.000	2708.	557311.	.7873	9163.3	1426.647
5	4.027	2581.	256998.	.5578	9163.3	1426.647
6	.891	2178.	-565484.	.1482	9242.8	1426.647
7	1.157	2280.	-366370.	.1833	9258.6	1426.647
8	1.124	2238.	-335211.	.1824	9221.4	1558.872
9	1.060	1626.	-1329661.	.2389	8712.6	1558.872
10	1.000	425.	-2807462.	.0000	.0	1558.872

Table 9.5 Continued

PATCHES 2/5/75 FOR PWRE FUNCT3 — DATE 022575 — PAGE

GENERATOR DESIGN

POINT	LENGTH	D/D4	P	T	SIGMA	H	RHO	U	MU	S	N	MU*B	K	B
400	.000	1.000	4.027	2501.	6.767	256998.	.5578	775.0	.2269	9163.3	.7291	1.3613	.820	6.00
401	.084	1.003	4.000	2579.	6.738	252963.	.5545	774.4	.2283	9163.7	.7776	1.3697	.819	6.00
402	.864	1.036	3.750	2561.	6.457	213386.	.5238	768.2	.2422	9166.6	.7760	1.4529	.814	6.00
403	1.668	1.073	3.500	2543.	6.165	171524.	.4929	761.7	.2579	9169.7	.7742	1.5475	.809	6.00
404	2.501	1.114	3.250	2522.	5.855	127063.	.4617	754.7	.2760	9173.1	.7722	1.6558	.804	6.00
405	3.369	1.160	3.000	2501.	5.536	79615.	.4303	747.1	.2969	9176.9	.7701	1.7813	.800	6.00
406	4.282	1.212	2.750	2477.	5.193	28667.	.3985	738.9	.3214	9181.0	.7678	1.9282	.795	6.00
407	5.249	1.271	2.500	2452.	4.838	-26394.	.3664	729.8	.3505	9185.6	.7655	2.1030	.790	6.00
408	6.286	1.341	2.250	2423.	4.456	-86233.	.3340	719.8	.3857	9190.8	.7629	2.3144	.785	6.00
409	7.503	1.423	2.000	2392.	4.057	-151821.	.3011	708.6	.4293	9196.8	.7570	2.3715	.780	5.52
410	9.091	1.523	1.750	2356.	3.630	-224303.	.2677	695.9	.4846	9203.8	.7508	2.4482	.775	5.05
411	11.245	1.647	1.500	2316.	3.172	-305779.	.2338	681.2	.5571	9212.3	.7444	2.5544	.770	4.58
412	15.052	1.846	1.260	2257.	2.580	-419705.	.1923	659.8	.6814	9225.1	.7366	2.7540	.765	4.04
413	16.842	1.930	1.100	2234.	2.368	-463071.	.1782	651.4	.7369	9230.2	.7335	2.8358	.763	3.85
414	19.017	2.026	1.000	2208.	2.149	-509760.	.1639	642.3	.8027	9235.9	.7303	2.9348	.761	3.66
415	21.712	2.139	.900	2180.	1.920	-560504.	.1495	632.1	.8821	9242.3	.7270	3.0562	.759	3.46
5	21.991	2.150	.891	2178.	1.898	-565484.	.1482	631.1	.8903	9242.8	.7266	3.0677	.759	3.45
6	.000	.000	1.157	2280.	3.060	-366370.	.1833	631.1	.7122	9258.6				

MASSFLOWS. GENERATOR AREA
MG 1426.647 MA 1248.159 MAW1256.134 MG*1558.872 MA* 132.224
MC 124.408 MC* 145.304 MS 25.213 MWC 20.895 MWS .000
(X + DH) * DH 71.10581 D4 1.8166 D5 3.9057 A4 3.3000 A515.2545
THETA * MC 4.03323+09 ETAMHD .76095

OVERALL PERFORMANCE

STEAM PLANT EFFICIENCY	.40	.41	.42	.43	.44	.45	.46
QS0	.23037+10	.23037+10	.23037+10	.23037+10	.23037+10	.23037+10	.23037+10
QS1	.20166+09	.20166+09	.20166+09	.20166+09	.20166+09	.20166+09	.20166+09
QS2	.11980+09	.11980+09	.11980+09	.11980+09	.11980+09	.11980+09	.11980+09
DWI	.17970+08	.17970+08	.17970+08	.17970+08	.17970+08	.17970+08	.17970+08
DWG	.12566+08	.12996+08	.13431+08	.13864+08	.14296+08	.14730+08	.15163+08
QS	.26424+10	.26428+10	.26433+10	.26437+10	.26441+10	.26446+10	.26450+10
PS	.10570+10	.10836+10	.11102+10	.11368+10	.11634+10	.11901+10	.12167+10
PT	.78535+09	.81238+09	.83942+09	.86647+09	.89353+09	.92060+09	.94768+09
PC	.28881+09	.28881+09	.20881+09	.28881+09	.28881+09	.28881+09	.28881+09
PSE	.77278+09	.79938+09	.82599+09	.85261+09	.87923+09	.90587+09	.93251+09
PMHD	.11800+10	.11800+10	.11800+10	.11800+10	.11800+10	.11800+10	.11800+10
PSE+PMHD	.19528+10	.19794+10	.20060+10	.20326+10	.20592+10	.20859+10	.21125+10
PAUX	.11592+08	.11991+08	.12390+08	.12789+08	.13189+08	.13588+08	.13988+08
P	.19412+10	.19674+10	.19936+10	.20198+10	.20460+10	.20723+10	.20985+10
HR	.70695+04	.69951+04	.69031+04	.68135+04	.67262+04	.66410+04	.65579+04
OVERALL EFFICIENCY	.4813	.4878	.4943	.5008	.5073	.5138	.5203

```
@FREE     ILL6WET*1PERCK295.

@ASG,A  ILL6DRY*1PERCK295.

@USE  4.,ILL6DRY*1PERCK295.

@XQT      MHDDUCT.ARECIRCINJOX
```

- The sorbent oxidizer produces some heat which can be used to generate steam.
- To obtain an efficiency (or heat rate) based on the raw coal, the ratio of fuel gas heating value to coal heating value was multiplied by the ratio of product fuel gas to coal.
- Part of the heat addition to the combustor is supplied by heating the fuel gas. In the calculation of fuel gas temperature, only sensible heat is considered (i.e., no change in composition of the fuel gas was considered).

Table 9.6 - Nomenclature for Additional Outputs for Open-Cycle MHD with Integrated Gasifier

AIR/FG	=	Ratio of flow rate of air to compressor to flow rate of fuel gas
QFG	=	Sensible heat addition to fuel gas in fuel heater, J/kg
TFUELGAS	=	Temperature of heated fuel gas, °K

Since only one coal (bituminous) and one process air temperature 672°K (750°F) were considered, most of the parameters (air/coal ratio, steam/coal ratio, heat from the sorbent oxidizer, etc.) were built into the program as constants.

9.3.3 Duct Program for Base Case 1

The use of a carbonizer and a separately fired preheater also required program modifications. The additional output quantities (added to those used for Base Case 2) are defined in Table 9.8. Table 9.9 is the output for Base Case 1 and the listing of the program (CHRDUC2) is in Appendix A 9.7.

The additional factors are accounted for in this program are that:

- Part of the air preheat is provided by the gapor so that all of the heat added to the air is not recovered from the MHD exhaust product.

Table 9.7 Output of DQHDUCT for Base Case 3

```
LOW-BTU GAS FROM ILL. 6-3% MOIS. WITH 1% K2CO3 SEED 750F AIR BLOWN
AIR WITH .639% MOIS. IS OXIDANT FUEL GAS IS SUPPLIED AT 1600F

PHI =    .95      PMHD =  1.37000+09      TCOMB =  2700.0      PCOMB =  10.000      UO =  775.00
RHRP =        .000      RHOX =        .000

RMAW =   5.643700-01        RMC =   4.233700-01        RMS =   1.225750-02
RMWC =    .000000          RMWS =    .000000          RMCA =   4.233700-01      RMAP =   5.940740-02

WC =   74.32        NC =    .9972        NA =    .9887      HHVDAF =   2.404900+03XS =   3.607440+00
TSTACK =    4.250000+02       TFUELGAS =   1.692000+03       AIR/FG =    6.682240-01

EFFICIENCIES
    DIFFUSER                        .8000
    ROTATING GENERATOR              .9840
    DC/AC INVERTER                  .9850

HEAT TRANSFER RATIO
    FROM COMBUSTER TO SUBPOSED PLANT     =      .050
    MHD GENERATOR TO SUBPOSED PLANT      =      .100

    FRICTION FACTOR IN MHD DUCT =     .0050
    PAUX/PSE =      .0150
    (PI-PCOMB)/PCOMB =      .0815
    PC/MAW =     4.74714+05
    QA/MAW =     1.32480+06
    THETA =          5.59375+06
    PMHD =         1.37000+09
    PE =           1.39086+09
    QFG =    6.40377+07

   C =      .1000
   DK =      .0157
   DB =     2.0000
   HG   1.258943+03
   HG   1.378472+03
   HG   1.374745+03
```

AIR SIDE	TOTAL PRESSURE	TEMPERATURE
COMPRESSOR INLET	.9800	288.3003
COMPRESSOR OUTLET	10.8150	591.2666
AIR PREHEATER EXIT	10.5000	1692.5629

GAS SIDE	TOTAL PRESSURE	TEMPERATURE
MHD DUCT INLET	10.0000	2700.0000
DIFFUSER EXIT	1.1573	2169.8205
INJECTOR EXIT	1.1236	2144.0210
AIR PREHEATER EXIT	1.0600	1465.3423
BOTTOMING HEAT EXCH. EXIT	1.0000	425.0000

GAS STATES

POINT,	P	T	H	RHO	S	M
0	1.000	288.	7.			775.933
1	.980	288.	7.			775.933
2	10.815	591.	140.			775.933
3	10.500	1693.	691.			775.933
4	10.000	2700.	169535.	1.2935	9181.9	1374.866
5	6.738	2572.	-130777.	.9199	9181.9	1374.866
6	.921	2078.	-1111645.	.1577	9329.6	1374.866
7	1.157	2170.	-943263.	.1894	9343.9	1374.866
8	1.129	2144.	-890317.	.1870	9323.5	1456.549
9	1.060	1465.	-1940406.	.2594	8739.1	1456.549
10	1.000	425.	-3233875.	.0000	.0	1456.544

Table 9.7 Continued

GENERATOR DESIGN

POINT	LENGTH	D/D4	P	T	SIGMA	H	RHO	U	MU	S	N	MU*B	K	B
400	.000	1.000	6.738	2572.	3.558	-130777.	.9199	775.0	.1393	9181.9	.6369	.8358	.820	6.00
401	1.208	1.019	6.500	2564.	3.510	-149193.	.8907	771.6	.1439	9185.1	.6750	.8636	.816	6.00
402	3.769	1.061	6.000	2544.	3.395	-191577.	.8290	763.9	.1548	9191.6	.6831	.9291	.809	6.00
403	6.403	1.109	5.500	2523.	3.261	-237794.	.7669	755.6	.1676	9198.4	.6900	1.0058	.801	6.00
404	7.765	1.136	5.250	2511.	3.189	-262596.	.7357	751.1	.1749	9201.9	.6930	1.0493	.797	6.00
405	9.153	1.164	5.000	2979.	3.115	-288574.	.7043	746.4	.1828	9205.6	.6958	1.0969	.794	6.00
406	10.572	1.195	4.750	2487.	3.033	-315860.	.6729	741.5	.1915	9209.3	.6983	1.1492	.790	6.00
407	12.026	1.229	4.500	2473.	2.946	-344618.	.6412	736.3	.2012	9213.3	.7005	1.2071	.786	6.00
408	13.520	1.265	4.250	2459.	2.856	-374940.	.6094	730.8	.2119	9217.4	.7026	1.2713	.782	6.00
409	15.061	1.305	4.000	2444.	2.761	-406954.	.5774	724.9	.2239	9221.8	.7044	1.3431	.778	6.00
410	16.655	1.349	3.750	2427.	2.658	-440859.	.5452	718.7	.2373	9226.4	.7060	1.4240	.774	6.00
411	18.314	1.397	3.500	2410.	2.550	-476920.	.5127	712.0	.2526	9231.2	.7074	1.5157	.771	6.00
412	20.045	1.452	3.250	2392.	2.436	-515405.	.4801	704.8	.2701	9236.4	.7086	1.6208	.767	6.00
513	21.866	1.512	3.000	2372.	2.311	-556668.	.4472	697.1	.2904	9242.0	.7096	1.7423	.763	6.00
414	23.793	1.581	2.750	2350.	2.183	-601156.	.4140	688.6	.3141	9248.1	.7105	1.8845	.759	6.00
415	25.851	1.661	2.500	2326.	2.039	-649402.	.3805	679.3	.3422	9254.6	.7111	2.0534	.755	6.00
416	28.074	1.753	2.250	2299.	1.890	-702195.	.3466	669.0	.3762	9261.9	.7116	2.2575	.752	6.00
417	30.694	1.863	2.000	2270.	1.725	-760160.	.3124	657.5	.4182	9270.1	.7058	2.3119	.748	5.53
418	34.122	1.996	1.750	2236.	1.553	-824332.	.2776	644.3	.4714	9279.7	.6999	2.3851	.744	5.06
419	38.764	2.163	1.500	2198.	1.369	-896361.	.2423	629.1	.5412	9291.2	.6937	2.4871	.741	4.60
420	46.954	2.429	1.200	2143.	1.124	-997038.	.1990	607.0	.6605	9308.3	.6861	2.6800	.736	4.06
421	50.784	2.543	1.100	2121.	1.035	-1035406.	.1844	598.2	.7137	9315.1	.6829	2.7590	.735	3.87
422	55.422	2.673	1.000	2098.	.947	-1076692.	.1695	588.6	.7769	9322.7	.6798	2.8548	.733	3.67
5	59.819	2.791	.921	2078.	.884	-1111695.	.1577	580.3	.8361	9329.6	.6778	2.9465	.732	3.52

MASSFLOWS, GENERATOR AREA

```
   MG 1374.866 MA   771.007 MAW 775.933 MG'1456.544 MA'   81.677
   MC  582.077 MCW 582.077 MS    16.852 MWC      .000 MWS     .000
    (X + DM) = DM 164.38871 C4 1.3886 DS 3.8759 A4 1.9283 A515.0225
    THETA = MC    3.80484+09 ETAMHD    .71436
```

OVERALL PERFORMANCE

STEAM PLANT EFFICIENCY	.40	.41	.42	.43	.44	.45	.46
QSU	.19439+10	.19439+10	.19439+10	.19439+10	.19439+10	.19439+10	.19439+10
QS1	.19024+09	.19024+09	.19024+09	.19024+09	.19024+09	.19024+09	.19024+09
QS2	.13909+09	.13909+09	.13909+09	.13909+09	.13909+09	.13909+09	.13909+09
DQI	.20863+08	.20863+08	.20863+08	.20863+08	.20863+08	.20863+08	.20863+08
DQG	.90113+07	.93864+07	.97617+07	.10137+08	.10513+08	.10888+08	.11264+08
QS	.22916+10	.22920+10	.22924+10	.22927+10	.22931+10	.22935+10	.22939+10
PS	.91665+09	.93972+09	.96280+09	.98588+09	.10090+10	.10321+10	.10552+10
PT	.56321+09	.58665+09	.61010+09	.63356+09	.65703+09	.68051+09	.70399+09
PC	.36835+09	.36835+09	.36835+09	.36835+09	.36835+09	.36835+09	.36835+09
PSE	.55420+09	.57727+09	.60034+09	.62343+09	.64652+09	.66962+09	.69273+09
PMHD	.13700+10	.13700+10	.13700+10	.13700+10	.13700+10	.13700+10	.13700+10
PSE+PMHD	.19242+10	.19473+10	.19703+10	.19934+10	.20165+10	.20396+10	.20627+10
PAUX	.83130+07	.86590+07	.90051+07	.93514+07	.96978+07	.10044+08	.10391+08
P	.19116+10	.19343+10	.19570+10	.19799+10	.20025+10	.20253+10	.20480+10
HR	.67917+04	.67119+04	.66339+04	.65577+04	.64832+04	.64104+04	.63392+04
OVERALL EFFICIENCY	.5024	.5084	.5144	.5203	.5263	.5323	.5383

```
2USE  4.,ILL6GAS.1PERCK295.

2XQT      MHDDUCT.DUCABS
```

- After providing the air preheat, there is still heat remaining in the products of combustion of the gapor which is used to preheat its combustion air and generate steam.
- There is additional heat input to the cycle (over and above that in the dry ash free combustible) in the form of gapor. This must be included in the calculation of efficiency or heat rate.
- Power is required to recirculate gapor products to limit flame temperatures to desirable values. This is discussed in Appendix A 9.5.

Table 9.8 - Nomenclature for Additional Outputs for Open-Cycle MHD with Carbonizer and Separately Fired Air Heater

HEXH = Specific enthalpy of products of combustion of gapor and air leaving separately fired air heater, J/kg

RMAMG = Ratio of combustion air for gapor to gapor flow rate

RMGMC = Ratio of gapor flow rate to ash free char flow rate

TAP = Preheat temperature of gapor combustion air, °K

TCR = Temperature of air to be heated at entrance to separately fired air preheater, °K

In addition to the changes in the program, a considerable amount of hand calculations were necessary to calculate:

- The temperature rise of the preheat stream for a given value of gapor combustion air preheat and temperature of the products leaving the separately fired air heaters. This involved iteration since the required ratio of recycled products (RMRP) changed when the temperature changed.

Table 9.9   Output of CHRDUC2 for Base Case 1

```
     CHAR PRODUCED BY CARBONIZATION OF ILL. #6 COAL, 1% POTASSIUM SEED
     OXIDANT IS AIR WITH 0.639% MOISTURE - HHV FROM JRH-CHAR +TAR

PHI =      .95       PTOT =  2.00000+09       TCOMB =  2700.0      PCOMB =   6.000      UO =  775.00
RHRP =          .300   RHOX =           .000

AIR HEATR.RECIRC.RATIO=1.8700 RATIO OF COAL TO CHAR HEAT VALUE= 1.2443

RMAR =    6.185942-01       RMC =    6.853927-02        RMS =   1.286656-02
RMWC =     .000000         RMWS =     .000000           RMCW =   6.853927-02      RMAP =    9.349331-02

WC =    16.03       NC =     .9979        NA =     .9861       HHVDAF =   1.363116+04
XS =    5.526298+00       T(STACK) =    4.250000+02

RMGMC =   5.129000-01       RMAMG =    5.586900+00       HEXH =  -1.268800+06       TAP=    1.580000+03       TCR =    1.580000+03
   MG  1.949161+03

JET PUMP EFFICIENCY= .150000

 EFFICIENCIES
    DIFFUSER                   .8000
    ROTATING GENERATOR         .9840
    DC/AC INVERTER             .9850

 HEAT TRANSFER RATIO
     FROM COMBUSTER TO SUBPOSED PLANT   =      .050
     MHD GENERATOR TO SUBPOSED PLANT    =      .100

     FRICTION FACTOR IN MHD DUCT =      .0050
     PAUX/PSE =      .0150
     (PI-PCOMB)/PCOMB =       .0815
     PC/MAR =     3.45060+05
     QA/MAR =    1.78745+06
     THETA =       3.94528+07
     PMHD =       1.16753+09
     PE =         1.18531+09

    C =     .1000
    DK =     .0208
    DB =    2.0000

AIR SIDE          TOTAL      PRESSURE      TEMPERATURE

   COMPRESSOR INLET            .9800         327.2519
   COMPRESSOR OUTLET          6.4890         559.4368
   AIR PREHEATER EXIT         6.3000        1885.3241

GAS SIDE                      TOTAL      PRESSURE      TEMPERATURE

   MHD DUCT INLET                          6.0000       2700.0000
   DIFFUSER EXIT                           1.1573       2267.0684
   INJECTOR EXIT                           1.1236       2242.3478
   AIR PREHEATER EXIT                      1.0600       1610.4935
   BOTTOMING HEAT EXCH. EXIT               1.0000        425.0000

         GAS STATES
```

POINT.	P	T	H	RHO	S	H
0	1.000	288.	7.			896.443
1	.980	327.	-549.			896.443
2	6.489	559.	-401.			1331.191
3	6.300	1885.	619.			1331.191
4	6.000	2700.	664394.	.8067	8951.7	1449.161
5	3.987	2568.	364382.	.5670	8951.7	1449.161
6	.881	2157.	-437137.	.1510	9029.4	1449.161
7	1.157	2267.	-235327.	.1883	9045.3	1449.161
8	1.124	2242.	-215122.	.1836	9130.4	1584.648
9	1.060	1610.	-1226290.	.2433	8611.9	1584.648
10	1.000	425.	-2666197.	.0000	.0	1584.648

Table 9.9 Continued

GENERATOR DESIGN

POINT	LENGTH	D/D4	P	T	SIGMA	H	RHO	U	MU	S	N	MU*B	K	B
400	.000	1.000	3.987	2568.	7.574	364082.	.5670	775.0	.2495	8951.7	.7315	1.4967	.820	6.00
401	.658	1.031	3.750	2552.	7.303	328699.	.5371	769.4	.2639	8954.8	.7798	1.5836	.815	6.00
402	1.374	1.067	3.500	2532.	6.956	287685.	.5055	763.0	.2812	8957.8	.7776	1.6875	.810	6.00
403	2.116	1.108	3.250	2512.	6.596	244121.	.4736	756.2	.3011	8961.1	.7753	1.8065	.805	6.00
404	2.891	1.153	3.000	2490.	6.220	197636.	.4414	748.8	.3241	8964.7	.7728	1.9444	.800	6.00
405	3.706	1.204	2.750	2466.	5.823	147750.	.4089	740.8	.3510	8968.6	.7702	2.1060	.795	6.00
406	4.571	1.263	2.500	2440.	5.408	93907.	.3761	732.0	.3831	8973.1	.7675	2.2983	.790	6.00
407	5.499	1.332	2.250	2411.	4.969	35454.	.3429	722.2	.4219	8978.1	.7647	2.5312	.785	6.00
408	6.592	1.413	2.000	2379.	4.505	-28593.	.3092	711.4	.4698	8983.8	.7586	2.5954	.780	5.52
409	8.022	1.512	1.750	2343.	4.016	-99412.	.2750	699.0	.5308	8990.7	.7523	2.6813	.775	5.05
410	9.968	1.634	1.500	2301.	3.493	-178894.	.2403	684.8	.6100	8998.9	.7458	2.7999	.770	4.58
411	13.426	1.830	1.200	2241.	2.815	-289935.	.1977	664.1	.7479	9011.4	.7377	3.0221	.765	4.04
412	15.062	1.912	1.100	2218.	2.572	-332165.	.1832	656.0	.8092	9016.3	.7345	3.1133	.763	3.85
413	17.057	2.007	1.000	2192.	2.325	-377682.	.1686	647.2	.8818	9021.9	.7312	3.2231	.761	3.65
414	19.538	2.117	.900	2163.	2.066	-427025.	.1539	637.3	.9696	9020.1	.7276	3.3581	.759	3.46
5	20.088	2.140	.881	2157.	2.015	-437137.	.1510	635.3	.9888	9029.4	.7268	3.3863	.758	3.42

MASSFLOWS, GENERATOR AREA
MG 1449.161 MA 890.751 MA* 896.443 MG*1584.648 MA* 135.487
MC 99.324 MCA 99.324 MS 18.646 MAC .000 MWS .000
(X + DM) * DM 65.40774 D4 1.8161 D5 3.8865 A4 3.2980 A515.1049
THETA * MC 3.91863+09 ETAMHD .76125

OVERALL PERFORMANCE

STEAM PLANT EFFICIENCY	.40	.41	.42	.43	.44	.45	.46
QSO	.22817+10	.22817+10	.22817+10	.22817+10	.22817+10	.22817+10	.22817+10
QS1	.19593+09	.19593+09	.19593+09	.19593+09	.19593+09	.19593+09	.19593+09
QS2	.11853+09	.11853+09	.11853+09	.11853+09	.11853+09	.11853+09	.11853+09
DQ1	.17513+08	.17513+08	.17513+08	.17513+08	.17513+08	.17513+08	.17513+08
DWG	.12044+08	.12471+08	.12899+08	.13327+08	.13755+08	.14183+08	.14612+08
QS	.26126+10	.26131+10	.26135+10	.26139+10	.26143+10	.26148+10	.26152+10
PS	.10451+10	.10719+10	.10977+10	.11240+10	.11503+10	.11766+10	.12030+10
PT	.75272+09	.77945+09	.80619+09	.83293+09	.85969+09	.88645+09	.91322+09
PC	.30933+09	.30933+09	.30933+09	.30933+09	.30933+09	.30933+09	.30933+09
PSE	.74068+09	.76698+09	.79329+09	.81960+09	.84593+09	.87227+09	.89861+09
PMHD	.11675+10	.11675+10	.11675+10	.11675+10	.11675+10	.11675+10	.11675+10
PSE+PMHD	.19082+10	.19345+10	.19608+10	.19871+10	.20135+10	.20398+10	.20661+10
PAUX	.11110+08	.11505+08	.11899+08	.12294+08	.12689+08	.13084+08	.13479+08
P	.18971+10	.19230+10	.19489+10	.19748+10	.20008+10	.20267+10	.20527+10
HR	.70981+04	.69532+04	.68607+04	.67707+04	.66829+04	.65974+04	.65140+04
OVERALL EFFICIENCY	.4841	.4907	.4973	.5040	.5106	.5172	.5238

- The enthalpy of the products leaving the separately fired air heater (HEXH).
- The quantity of gapor products to be recirculated to limit the flame temperature to desirable values. These hand calculations are described in Appendix A 9.8.

9.3.4 Modifications to Duct Program Calculations

To meet the time schedule of the project, equipment calculations were made in parallel with the thermodynamic calculations of the duct program. As a result, the energy and material requirements of these are not included in the duct program outputs but are accounted for in the overall economic program as auxiliary powers or additional fuel inputs.

The major external energy consumer is the seed treatment plant (Appendix A 9.1). For Base Cases 1 and 2 and their variations the potassium seed is used to remove sulfur. It is necessary, therefore, to convert enough of the collected potassium sulfate to potassium carbonate to react with the sulfur from the fuel. This is a complex chemical engineering problem which requires power and fuel input and returns some fuel to the plant. An adjustment to the coal flow calculated from the duct program is made in terms of an equivalent coal flow and auxiliary power required by the seed treatment plant.

The precipitators used in all cases also require some power input which is subtracted from the power output calculated by the duct program. Other powers required by auxiliary equipment (such as cooling towers or coal crushers) are subjected in the overall economic program.

9.4 Results of Parametric Study

As in the discussion of parametric points to be investigated, the results will be discussed in order of increasing cycle complexity (i.e., Base Cases 2, 3, and 1).

9.4.1 Base Case 2 and Variations

The detailed accounts listing, the cost of electricity summary, and the input-output sheet for Base Case 2 are included as Tables 9.10, 9.11, and 9.12, respectively. Table 9.10 shows that 66.3% of the direct cost of the plant lies in Accounts 11, 12, 13, and 15. The MHD duct direct cost (Subaccounts 11.1 to 11.4) was calculated to be about \$1,486,000, which is small compared to the cost of the superconducting magnet (about \$7,800,000); and steam turbine-driven compressor (\$16,975,000); the steam turbine generator (\$27,570,000); heat recovery steam generator (\$58,518,000); the inverter-filter system (\$65,020,000); the seed treatment plant (\$27,830,000); and the air heaters and stoves (\$147,110,000). With the exception of the steam turbine-generator and steam turbine-compressor, and the possible exception of the inverter system, the remaining items are unproved for this duty or have never been built. For the purpose of this study they have been assumed to be fully developed and to have the required 30-year life. The conceptual design and sizing of these subsystems in the appendices of this section have resulted in material requirement descriptions for each of these components. Direct costs were generated from these descriptions.

Table 9.11 shows the importance of the field labor rate, the contingency allowance, the escalation rate, the rate of interest during construction, the fixed charge rate, the fuel cost, and the capacity factor on the cost of electricity. For a capital intensive plant such as this the importance of the capacity factor is clear, as is the need to base load the plant.

Table 9.12 shows the additional auxiliary power which was deducted from the nominal power of the station to arrive at the net station power (in this case 55.39 MWe or 3.28% of the nominal station power).

The fuel for the base case is bituminous coal with 13% moisture (as received). The subtitle "Points" in the listing titled "Gas States" at the bottom of Table 9.5 refers to the number on the schematic diagram

Table 9.10 BC NO 2 OPEN CYCLE MHD-STEAM BOTTOMING ACCOUNT LISTING
PARAMETRIC POINT NO. 1

ACCOUNT NO. & NAME,	UNIT	AMOUNT	MAT $/UNIT	INS $/UNIT	MAT COST,$	INS COST,$
SITE DEVELOPMENT						
1. 1 LAND COST	ACRE	232.0	1000.00	.00	232000.00	.00
1. 2 CLEARING LAND	ACRE	77.3	.00	600.00	.00	46395.36
1. 3 GRADING LAND	ACRE	232.0	.00	3000.00	.00	696000.00
1. 4 ACCESS RAILROAD	MILE	5.0	115000.00	110000.00	575000.00	550000.00
1. 5 LOOP RAILROAD TRACK	MILE	3.0	120000.00	70000.00	360000.00	210000.00
1. 6 SIDING R R TRACK	MILE	.0	125000.00	80000.00	.00	.00
1. 7 OTHER SITE COSTS	ACRE	.0	.00	.00	479082.76	479082.76
PERCENT TOTAL DIRECT COST IN ACCOUNT 1 = .633 ACCOUNT TOTAL,$					1646082.75	1981478.09
EXCAVATION & PILING						
2. 1 COMMON EXCAVATION	YD3	200100.0	.00	3.00	.00	600300.00
2. 2 PILING	FT	533600.0	6.50	8.50	3468400.00	4535600.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 2 = 1.430 ACCOUNT TOTAL,$					3468400.00	5135900.00
PLANT ISLAND CONCRETE						
3. 1 PLANT IS. CONCRETE	YD3	66700.0	70.00	80.00	4669000.00	5336000.00
3. 2 SPECIAL STRUCTURES	YD3	.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 3 = 1.662 ACCOUNT TOTAL,$					4669000.00	5336000.00
HEAT REJECTION SYSTEM						
4. 1 COOLING TOWERS	EACH	28.0	.00	.00	4298000.00	2142000.00
4. 2 CIRCULATING H2O SYS	EACH	1.0	.00	.00	1845746.97	2474913.41
4. 3 SURFACE CONDENSER	FT2	673513.8	.00	.00	3051513.50	471459.69
PERCENT TOTAL DIRECT COST IN ACCOUNT 4 = 2.373 ACCOUNT TOTAL,$					9195260.37	5088373.06
STRUCTURAL FEATURES						
5. 1 STAT. STRUCTURAL ST.	TON	3350.0	650.00	175.00	2177500.00	586250.00
5. 2 SILOS & BUNKERS	TPH	637.9	1800.00	750.00	1148177.50	478407.30
5. 3 CHIMNEY	FT	400.0	.00	.00	435070.92	652606.38
5. 4 STRUCTURAL FEATURES	EACH	1.0	966000.00	231000.00	966000.00	231000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 5 = 1.109 ACCOUNT TOTAL,$					4726748.37	1948263.67
BUILDINGS						
6. 1 STATION BUILDINGS	FT3	8375000.0	.16	.16	1340000.00	1340000.00
6. 2 ADMINISTRATION	FT2	15000.0	16.00	14.00	240000.00	210000.00
6. 3 WAREHOUSE & SHOP	FT2	24000.0	12.00	8.00	288000.00	192000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 6 = .600 ACCOUNT TOTAL,$					1868000.00	1742000.00
FUEL HANDLING & STORAGE						
7. 1 COAL HANDLING SYS	TPH	672.8	.00	.00	11980123.37	4772939.94
7. 2 DOLOMITE HAND. SYS	TPH	18.5	.00	.00	350391.39	222750.91
7. 3 FUEL OIL HAND. SYS	GAL	2600000.0	.30	.00	290836.01	227826.41
PERCENT TOTAL DIRECT COST IN ACCOUNT 7 = 2.965 ACCOUNT TOTAL,$					12621350.75	5223517.19
FUEL PROCESSING						
8. 1 COAL DRYER & CRUSHER	TPH	637.9	.00	.00	1961329.23	1307552.83
8. 2 CARBONIZERS	TPH	.0	.00	.00	.00	.00
8. 3 GASIFIERS	TPH	.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 8 = .543 ACCOUNT TOTAL,$					1961329.23	1307552.83

Table 9.10 Continued

.9C NO 2 OPEN CYCLE MHD-STEAM BOTTOMING ACCOUNT LISTING
PARAMETRIC POINT NO. 1

ACCOUNT NO. & NAME.	UNIT	AMOUNT	MAT $/UNIT	INS $/UNIT	MAT COST,$	INS COST,$
FIRING SYSTEM						
9. 1 CONTAINMENT STEEL	KP	479.4	350.00	159.00	167790.00	76224.60
9. 2 BRICK- DENSE ZRO2	KP	108.1	3305.00	150.00	357270.50	16215.00
9. 3 BRICK-SILICON CARBIDE	KP	.0	1500.00	159.00	.00	.00
9. 4 BRICK- MGO	KP	72.7	131.00	150.00	9523.70	10905.00
9. 5 BRICK (INSULATING)	KP	23.0	159.00	150.00	3657.00	3450.00
9. 6 STRUCTURAL STEEL	KP	204.9	509.00	156.00	104294.10	31964.40
PERCENT TOTAL DIRECT COST IN ACCOUNT 9 =			.130	ACCOUNT TOTAL,$	642535.29	138759.00
VAPOR GENERATOR (FIRED)						
10. 1		.0	.30	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 10 =			.000	ACCOUNT TOTAL,$	.00	.00
ENERGY CONVERTER						
11. 1 DUCT INSULATION BRICK	KP	19.8	137.00	274.00	2712.60	5425.20
11. 2 DUCT CONDUCTING BRICK	KP	19.8	500.00	1225.00	9900.00	24255.00
11. 3 DUCT STRUCTURAL STEEL	KP	137.9	3000.00	3000.00	413700.00	413700.00
11. 4 DUCT COOLING TUBES	KP	102.7	3000.00	3000.00	308100.00	308100.00
11. 5 SUPERCONDUCTING MAT	EACH	1.0	31999999.75	7999999.94	31999999.75	7999999.94
11. 6 MAGNET STRUCT STEEL	EACH	1.0	37799999.50	.00	37799999.50	.00
11. 7 NITROGEN LIQUIFIER	EACH	1.0	81000.00	9000.00	81000.00	9000.00
11. 8 HELIUM LIQUIFIER	EACH	1.0	270000.00	30000.00	270000.00	30000.00
11. 9 COMPRESSOR	EACH	1.0	5600000.00	559999.99	5600000.00	559999.99
11.10 COMPRESSOR DRIVE	EACH	1.0	10000000.00	815132.06	10000000.00	815132.06
11.11 STEAM TURB-GEN	EACH	1.0	26000000.00	1570121.95	26000000.00	1570121.95
11.12 FEED WATER HEATERS	PLANT	1.0	580000.00	1740000.00	580000.00	1740000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 11 =			21.024	ACCOUNT TOTAL,$	113065409.00	13475733.62
COUPLING HEAT EXCHANGER						
12. 1 HIGH TEMP SUPERHEATER	EA	1.0	3520799.97	1760399.98	3520799.97	1760399.98
12. 2 REHEATER & EVAP SECT	EA	1.0	9885600.00	6590399.94	9885600.00	6590399.94
12. 3 ECONOMIZER SECTION	EA	1.0	25732392.00	11028168.00	25732392.00	11028168.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 12 =			9.722	ACCOUNT TOTAL,$	39138791.50	19378967.75

Table 9.10 3C NO 2 OPEN CYCLE MHD-STEAM BOTTOMING ACCOUNT LISTING
Continued
PARAMETRIC POINT NO. 1

ACCOUNT NO. & NAME,	UNIT	AMOUNT	MAT $/UNIT	INS $/UNIT	MAT COST,$	INS COST,$
HEAT RECOVERY HEAT EXCH.						
13. 1 CERAMIC TUBING	KP	635.0	8640.00	1020.00	5486400.00	647700.00
13. 2 EXOTIC METAL TUBES	KP	5913.0	11316.00	6720.00	55745960.00	39043200.00
13. 3 STAINLESS STEEL TUBES	KP	6293.0	3000.00	1800.00	18879000.00	11327400.00
13. 4 TUBE CERAMIC COATING	KP	157.0	585.60	513.60	91939.20	80635.20
13. 5 INSULATION FOR REGEN	KP	4086.0	230.40	180.00	941414.40	735480.00
13. 6 STRUCTURAL STEEL	KP	190.0	564.00	187.20	107160.00	35568.00
13. 7 CONT STEEL REGEN	KP	1938.0	420.00	190.80	813960.00	369770.40
13. 8 CHECKER BRICKS	KP	.0	225.60	180.00	.00	.00
13. 9 INSULATION SFA HEATER	KP	.0	230.40	180.00	.00	.00
13.10 CONT ST SFA HEATER	KP	.0	420.00	190.80	.00	.00
13.11 STRU ST SFA HEATER	KP	.0	564.00	187.20	.00	.00
13.12 BURNER SFA HEATER	EACH	.0	.00	.00	.00	.00
13.13 HIGH TEMP VALVES	EACH	.0	.00	.00	.00	.00
13.14 PREHEATER AIR PIPING	TON	.0	.00	.00	.00	.00
13.15 FUEL GAS PIPING	TON	.0	.00	.00	.00	.00
13.16 COMB AIR PIPING	TON	.0	.00	.00	.00	.00
13.17 RECIRC AIR PIPING	TON	.0	.00	.00	.00	.00
13.18 PRODUCTS PIPING	TON	.0	.00	.00	.00	.00
13.19 RECIRC ASPIRATOR	TON	.0	.00	.00	.00	.00
13.20 HEADERS	KP	645.0	912.00	2136.00	588240.00	1377720.00
13.21 SEALS	EACH	7200.0	20.40	40.80	146880.00	293760.00
13.22 CONCRETE	YD3	1903.0	84.00	96.00	159852.00	182688.00
13.23 EXPANSION JOINTS	EACH	1.0	48000.00	12000.00	48000.00	12000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 13 =24.442 ACCOUNT TOTAL,$					93008804.00	54105920.00
WATER TREATMENT						
14. 1 DEMINERALIZER	GPM	368.9	2000.00	560.00	737872.68	206604.35
14. 2 CONDENSATE POLISHING	KWE	1167500.0	1.25	.30	1459374.97	350250.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 14 = .458 ACCOUNT TOTAL,$					2197247.62	556854.34
POWER CONDITIONING						
15. 1 INVERTER SYSTEM	KWE	1190000.0	43.30	5.37	50740000.50	6342500.06
15. 2 FILTERS	KWE	1180000.0	6.00	.73	7080000.06	855500.01
15. 3 STD TRANSFORMERS	KWE	1073966.7	.00	.00	1828159.62	36563.19
PERCENT TOTAL DIRECT COST IN ACCOUNT 15 =11.112 ACCOUNT TOTAL,$					59648160.00	7234563.25
AUXILIARY MECH EQUIPMENT						
16. 1 BOILER FEED PUMP &DR.	KWE	1109125.0	1.67	.10	1852238.72	110912.50
16. 2 OTHER PUMPS	KWE	977882.5	.88	.12	860536.60	117345.90
16. 3 MISC SERVICE SYS	KWE	1955765.0	1.17	.73	2288245.03	1427708.44
16. 4 AUXILIARY BOILER	PPH	480000.0	4.00	.80	1920000.00	384000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 16 = 1.489 ACCOUNT TOTAL,$					6921020.31	2039966.81
PIPE & FITTINGS						
17. 1 CONVENTIONAL PIPING	TON	3800.0	3000.00	1800.00	11400000.00	6840000.00
17. 2 HIGT TEMP AIR PIPING	TON	3398.0	450.00	225.00	1529100.00	764550.00
17. 3 LOW TEMP AIR PIPING	TON	107.0	1200.00	800.00	128400.00	85600.00
17. 4 RECIR PRODUCT PIPING	TON	.0	1200.00	800.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 17 = 3.447 ACCOUNT TOTAL,$					13057500.00	7690150.00

Table 9.10 Continued

3C NO 2 OPEN CYCLE MHD-STEAM BOTTOMING ACCOUNT LISTING
PARAMETRIC POINT NO. 1

ACCOUNT NO. & NAME,	UNIT	AMOUNT	MAT $/UNIT	INS $/UNIT	MAT COST,$	INS COST,$
AUXILIARY ELEC EQUIPMENT						
18. 1 MISC MOTERS,ETC		1564612.0	1.40	.17	2190456.78	265984.04
18. 2 SWITCHGEAR & MCC PAN	KWE	1564612.0	1.95	.45	3050993.34	704075.39
18. 3 CONDUIT,CABLES,TRAYS	FT	6476000.0	1.32	1.36	8548319.87	8807359.87
18. 4 ISOLATED PHASE BUS	FT	570.0	510.00	450.00	290700.00	256500.00
18. 5 LIGHTING & COMMUN	KWE	1955765.0	.35	.43	684517.75	840978.95
PERCENT TOTAL DIRECT COST IN ACCOUNT 18 = 4.260 ACCOUNT TOTAL,$					14764987.75	10874898.12
CONTROL, INSTRUMENTATION						
19. 1 COMPUTER	EACH	1.0	660000.00	15000.00	660000.00	15000.00
19. 2 OTHER CONTROLS	EACH	1.0	1290000.00	774000.00	1290000.00	774000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 19 = .455 ACCOUNT TOTAL,$					1950000.00	789000.00
PROCESS WASTE SYSTEMS						
20. 1 BOTTOM ASH	TPH	51.7	4314704.87	1228675.22	4914704.87	1228675.22
20. 2 DRY ASH	TPH	12.9	899053.76	224763.44	899053.76	224763.44
20. 3 WET SLURRY	TPH	.0	.00	.00	.00	.00
20. 4 ONSITE DISPOSAL	ACRE	281.4	6456.33	9594.00	1816733.41	2699638.44
20. 5 SEED TREATMENT	EACH	1.0	18546100.00	9183899.87	18646100.00	9183899.87
PERCENT TOTAL DIRECT COST IN ACCOUNT 20 = 6.581 ACCOUNT TOTAL,$					26276592.00	13336977.87
STACK GAS CLEANING						
21. 1 PRECIPITATOR	EACH	1.0	25900000.00	3540000.00	25900000.00	3540000.00
21. 2 SCRUBBER	KWE	.0	16.34	7.05	.00	.00
21. 3 MISC STEEL & DUCTS		1.0	4250000.00	.00	4250000.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 21 = 5.597 ACCOUNT TOTAL,$					30150000.00	3540000.00
TOTAL DIRECT COSTS,$					440977208.00	160924870.00

Table 9.11

BC NO 2 OPEN CYCLE MHD-STEAM BOTTOMING COST OF ELECTRICITY,MILLS/KW.HR

PARAMETRIC POINT NO. 1

ACCOUNT	RATE, PERCENT	LABOR RATE, $/HR 6.00	8.50	10.60	15.00	21.50
TOTAL DIRECT COSTS,$	.0	532066755.	570020728.	601902072.	658701072.	767381424.
INDIRECT COST,$	51.0	46455669.	65812198.	82071682.	116139172.	166466148.
PROF & OWNER COSTS,$	8.0	42565340.	45601658.	48152165.	53496085.	61390513.
CONTINGENCY COST,$	10.0	53206675.	57002072.	60190206.	66870106.	76738142.
SUB TOTAL,$	.0	674294432.	738436648.	792316112.	905206424.	1071976216.
ESCALATION COST,$	6.5	202104614.	221329802.	237478962.	271315296.	321300796.
INTREST DURING CONST,$	10.0	249027099.	272715784.	292614284.	334306376.	395896976.
TOTAL CAPITALIZATION,$	.0	1125426128.	1232482224.	1322409344.	1510828080.	1789173984.
COST OF ELEC-CAPITAL	18.0	17.84833	19.54615	20.97232	23.96048	28.37482
COST OF ELEC-FUEL	.0	6.19012	6.19012	6.19012	6.19012	6.19012
COST OF ELEC-OP & MAIN	.0	.84737	.84737	.84737	.84737	.84737
TOTAL COST OF ELEC	.0	24.88582	26.58364	28.00981	30.99798	35.41231

ACCOUNT	RATE, PERCENT	CONTINGENCY, PERCENT -5.00	.00	10.00	5.00	20.00
TOTAL DIRECT COSTS,$	.0	601902072.	601902072.	601902072.	601902072.	601902072.
INDIRECT COST,$	51.0	82071682.	82071682.	82071682.	82071682.	82071682.
PROF & OWNER COSTS,$	8.0	48152165.	48152165.	48152165.	48152165.	48152165.
CONTINGENCY COST,$	20.0	-30095103.	0.	60190206.	30095103.	120380413.
SUB TOTAL,$	.0	702030816.	732125912.	792316112.	762221008.	852506320.
ESCALATION COST,$	6.5	210417972.	219438300.	237478962.	228458630.	255519624.
INTREST DURING CONST,$	10.0	259270564.	270385136.	292614284.	281499708.	314843436.
TOTAL CAPITALIZATION,$	.0	1171719344.	1221949344.	1322409344.	1272179328.	1422869360.
COST OF ELEC-CAPITAL	18.0	18.58250	19.37910	20.97232	20.17571	22.56553
COST OF ELEC-FUEL	.0	6.19012	6.19012	6.19012	6.19012	6.19012
COST OF ELEC-OP & MAIN	.0	.84737	.84737	.84737	.84737	.84737
TOTAL COST OF ELEC	.0	25.62000	26.41660	28.00981	27.21321	29.60303

ACCOUNT	RATE, PERCENT	ESCALATION RATE, PERCENT 5.00	6.50	8.00	10.00	.00
TOTAL DIRECT COSTS,$	.0	601902072.	601902072.	601902072.	601902072.	601902072.
INDIRECT COST,$	51.0	82071682.	82071682.	82071682.	82071682.	82071682.
PROF & OWNER COSTS,$	8.0	48152165.	48152165.	48152165.	48152165.	48152165.
CONTINGENCY COST,$	10.0	60190206.	60190206.	60190206.	60190206.	60190206.
SUB TOTAL,$	.0	792316112.	792316112.	792316112.	792316112.	792316112.
ESCALATION COST,$	.0	177805296.	237478962.	300304688.	389211164.	0.
INTREST DURING CONST,$	10.0	278369680.	292614284.	307450504.	328219632.	235133662.
TOTAL CAPITALIZATION,$	.0	1248511088.	1322409344.	1400071296.	1509746896.	1027449768.
COST OF ELEC-CAPITAL	18.0	19.80035	20.97232	22.20397	23.94334	16.29450
COST OF ELEC-FUEL	.0	6.19012	6.19012	6.19012	6.19012	6.19012
COST OF ELEC-OP & MAIN	.0	.84737	.84737	.84737	.84737	.84737
TOTAL COST OF ELEC	.0	26.83785	28.00981	29.24147	30.98083	23.33200

ACCOUNT	RATE, PERCENT	INT DURING CONST,PERCENT 6.00	8.00	10.00	12.50	15.00
TOTAL DIRECT COSTS,$	.0	601902072.	601902072.	601902072.	601902072.	601902072.
INDIRECT COST,$	51.0	82071682.	82071682.	82071682.	82071692.	82071682.
PROF & OWNER COSTS,$	8.0	48152165.	48152165.	48152165.	48152165.	48152165.
CONTINGENCY COST,$	10.0	60190206.	60190206.	60190206.	60190206.	60190206.
SUB TOTAL,$	.0	792316112.	792316112.	792316112.	792316112.	792316112.
ESCALATION COST,$	6.5	237478962.	237478962.	237478962.	237478962.	237478962.
INTREST DURING CONST,$	15.0	167843940.	228875416.	292614284.	376254776.	464500600.
TOTAL CAPITALIZATION,$	.0	1197639008.	1258670480.	1322409344.	1406049840.	1494295664.
COST OF ELEC-CAPITAL	18.0	18.99356	19.96147	20.97232	22.29879	23.69829
COST OF ELEC-FUEL	.0	6.19012	6.19012	6.19012	6.19012	6.19012
COST OF ELEC-OP & MAIN	.0	.84737	.84737	.84737	.84737	.84737
TOTAL COST OF ELEC	.0	26.03106	26.99897	28.00981	29.33628	30.73579

Table 9.11 Continued -

BC NO 2 OPEN CYCLE MHD-STEAM BOTTOMINC COST OF ELECTRICITY,MILLS/KW.HR
PARAMETRIC POINT NO. 1

ACCOUNT	RATE, PERCENT	FIXED CHARGE RATE, PCT 10.00	14.40	18.00	21.60	25.00
TOTAL DIRECT COSTS,$	.0	601902072.	601902072.	601902072.	601902072.	601902072.
INDIRECT COST,$	51.0	82071682.	82071682.	82071682.	82071682.	82071682.
PROF & OWNER COSTS,$	9.0	48152165.	48152165.	48152165.	48152165.	48152165.
CONTINGENCY COST,$	10.0	60190206.	60190206.	60190206.	60190206.	60190206.
SUB TOTAL,$	.0	792316112.	792316112.	792316112.	792316112.	792316112.
ESCALATION COST,$	6.5	237478962.	237478962.	237478962.	237478962.	237478962.
INTREST DURING CONST,$	10.0	292614284.	292614284.	292614284.	292614284.	292614284.
TOTAL CAPITALIZATION,$	.0	1322409344.	1322409344.	1322409344.	1322409344.	1322409344.
COST OF ELEC-CAPITAL	25.0	11.65129	16.77785	20.97232	25.16678	29.12822
COST OF ELEC-FUEL	.0	6.19012	6.19012	6.19012	6.19012	6.19012
COST OF ELEC-OP & MAIN	.0	.84737	.84737	.84737	.84737	.84737
TOTAL COST OF ELEC	.0	18.68878	23.81535	28.00981	32.20428	36.16572

ACCOUNT	RATE, PERCENT	FUEL COST, $/10**6 BTU .50	.85	1.50	2.50	1.02
TOTAL DIRECT COSTS,$	.0	601902072.	601902072.	601902072.	601902072.	601902072.
INDIRECT COST,$	51.0	82071682.	82071682.	82071682.	82071682.	82071682.
PROF & OWNER COSTS,$	8.0	48152165.	48152165.	48152165.	48152165.	48152165.
CONTINGENCY COST,$	10.0	60190205.	60190206.	60190206.	60190206.	60190206.
SUB TOTAL,$	.0	792316112.	792316112.	792316112.	792316112.	792316112.
ESCALATION COST,$	5.5	237478962.	237478962.	237478962.	237478962.	237478962.
INTREST DURING CONST,$	10.0	292614284.	292614284.	292614284.	292614284.	292614284.
TOTAL CAPITALIZATION,$	.0	1322409344.	1322409344.	1322409344.	1322409344.	1322409344.
COST OF ELEC-CAPITAL	18.0	20.97232	20.97232	20.97232	20.97232	20.97232
COST OF ELEC-FUEL	.0	3.64125	6.19012	10.92375	18.20625	7.42815
COST OF ELEC-OP & MAIN	.0	.84737	.84737	.84737	.84737	.84737
TOTAL COST OF ELEC	.0	25.46094	28.00981	32.74344	40.02594	29.24784

ACCOUNT	RATE, PERCENT	CAPACITY FACTOR, PERCENT 12.00	45.00	50.00	65.00	80.00
TOTAL DIRECT COSTS,$	.0	501902072.	601902072.	601902072.	601902072.	501902072.
INDIRECT COST,$	51.0	82071682.	82071682.	82071682.	82071682.	82071682.
PROF & OWNER COSTS,$	8.0	48152165.	48152165.	48152165.	48152165.	48152165.
CONTINGENCY COST,$	10.0	60190206.	60190206.	60190206.	60190206.	60190206.
SUB TOTAL,$	.0	792316112.	792316112.	792316112.	792316112.	792316112.
ESCALATION COST,$	6.5	237478962.	237478962.	237478962.	237478962.	237478962.
INTREST DURING CONST,$	10.0	292614284.	292614284.	292614284.	292614284.	292614284.
TOTAL CAPITALIZATION,$	.0	1322409344.	1322409344.	1322409344.	1322409344.	1322409344.
COST OF ELEC-CAPITAL	18.0	113.60005	30.29335	27.26401	20.97232	17.04001
COST OF ELEC-FUEL	.0	6.19012	6.19012	6.19012	6.19012	6.19012
COST OF ELEC-OP & MAIN	.0	.84738	.84737	.84737	.84737	.84737
TOTAL COST OF ELEC	.0	120.63756	37.33084	34.30151	28.00981	24.07750

Table 9.12 BC NO 2 OPEN CYCLE MHD-STEAM BOTTOMING

ACCOUNT NO	AUX POWER,MWE	PERC PLANT POW	OPERATION COST	MAINTENANCE COST
4	15.70249	.78776	89.02623	24.45828
7	6.42215	.32219	.00000	.00000
8	.92924	.04160	.00000	.00000
13	.00000	.00000	13.21918	.00000
14	.00000	.00000	33.92728	.00000
18	15.56570	.78090	.00000	.00000
20	24.41190	1.22459	1113.18802	.00000
21	2.46000	.12341	.00000	.00000
TOTALS	65.39146	3.28055	1689.07141	24.45828

BC NO 2 OPEN CYCLE MHD-STEAM BOTTOMING BASE CASE INPUT

NOMINAL POWER, MWE	2058.7000	NET POWER, MWE	1993.3085
NOM HEAT RATE, BTU/KW-HR	7051.1821	NET HEAT RATE, BTU/KW-HR	7282.4994
OFF DESIGN HEAT RATE	1.0367		
CONDENSER			
DESIGN PRESSURE, IN HG A	2.0000	NUMBER OF SHELLS	5.0000
NUMBER OF TUBES/SHELL	6839.0271	TUBE LENGTH, FT	75.2339
U, BTU/HR-FT2-F	609.8535	TERMINAL TEMP DIFF, F	5.0000
HEAT REJECTION			
DESIGN TEMP, F	51.4000	APPROACH, F	21.6744
RANGE, F	23.0000	OFF DESIGN TEMP, F	77.0000
OFF DESIGN PRES, IN HG A	2.9527	LP TURBINE BLADE LEN, IN	31.5000

1	2058.700	2	.000	3	.510	4	6685.200	5	7.000
6	879.700	7	2.000	8	1612.200	9	5.000	10	3.000
11	1.000	12	288.800	13	1.000	14	.000	15	.000
16	2.000	17	232.000	18	3.000	19	5.000	20	3.000
21	.000	22	66700.000	23	.000	24	3350.000	25	400.000
26	8375000.000	27	15000.000	28	24000.000	29	2600000.000	30	.500
31	1.000	32	3800.000	33	1.000	34	.800	35	.800
36	6475000.000	37	570.000	38	1.000	39	1.000	40	966000.000
41	231000.000	42	660000.000	43	15000.000	44	1290000.000	45	774000.000
46	6.000	47	.000	48	3.000	49	1.000	50	2.000
51	288.800	52	5.350						

1	479.400	2	108.100	3	.000	4	72.700	5	23.000
6	204.900	7	350.000	8	159.000	9	3305.000	10	150.000
11	1500.000	12	159.000	13	131.000	14	150.000	15	159.000
16	150.000	17	509.000	18	156.000	19	19.800	20	19.800
21	137.900	22	102.700	23	1.000	24	1.000	25	1.000
26	1.000	27	1.000	28	1.000	29	1.000	30	1.000
31	137.000	32	274.000	33	500.000	34	1225.000	35	3000.000
36	3000.000	37	3000.000	38	3000.000	39	40000000.000	40	.000
41	42000000.073	42	.000	43	90000.000	44	.000	45	300000.000
46	.000	47	5600060.000	48	.000	49	10000000.000	50	.000
51	26000000.000	52	.000	53	580000.000	54	3.000	55	.000
56	1.000	57	1.000	58	1.000	59	3260000.000	60	.000
61	13730000.000	62	.300	63	17210000.000	64	.000	65	480000.000
66	1.000	67	.000	68	.000	69	.000	70	.000
71	.000	72	.000	73	.000	74	.000	75	.000
76	.000	77	.000	78	.000	79	635.000	80	5810.000
81	5293.000	82	157.000	83	4086.000	84	190.000	85	1938.000
86	.000	87	.000	88	.000	89	.000	90	.000
91	.000	92	.000	93	.000	94	.000	95	.000
96	.000	97	.000	98	645.000	99	7200.000	100	1903.000
101	1.000	102	9640.000	103	1020.000	104	11316.000	105	6720.000
106	3000.000	107	1800.000	108	585.600	109	513.600	110	230.400
111	190.000	112	564.000	113	187.200	114	420.000	115	190.800
116	225.600	117	180.000	118	230.400	119	180.000	120	420.000
121	190.800	122	564.000	123	187.200	124	.000	125	.000
126	.000	127	.000	128	.000	129	.000	130	.000
131	.000	132	.000	133	.000	134	.000	135	.000
136	.000	137	.000	138	.000	139	.000	140	.000
141	912.000	142	2135.000	143	23.400	144	40.800	145	84.000
146	96.000	147	48000.000	148	12000.000	149	43.000	150	21.500
151	6.000	152	2.900	153	3398.000	154	107.000	155	.000
156	450.000	157	225.000	158	1200.000	159	800.000	160	1200.000
161	800.000	162	1.000	163	27930000.000	164	.000	165	1.000
166	.000	167	1.000	168	25900000.000	169	3540000.000	170	.000
171	.000	172	4250000.000	173	.000	174	16.790	175	2.460
176	8.800	177	.133	178	1.054	179	8.298	180	11.470
181	1.100	182	.000						

(Figure 9.1). These data describe the gas thermodynamic state and mass flow rate at 10 points in the cycle and are in metric units. The combustor pressure is 6.078 MPa (6 atm), and with 1588°K (2398°F) preheat temperature the combustor temperature is 2708°K (4414°F). The overall plant efficiency is about 47%. Due to the relatively high sulfur content of this coal, it is necessary to convert nearly all of the potassium sulfate to potassium carbonate in the seed treatment plant. The energy and power requirements of the seed treatment (Appendix A 9.1) reduce the efficiency approximately three points. The capital cost of the plant (including escalation, interest during construction, and contingency) is \$663/kW for a 1993 MWe plant. The O&M charge is about (0.236 mills/MJ (0.85 mills/kWh). This is partly manpower and makeup water, but largely [0.139 mills/MJ/kWh ($\sim$ 0.5 mills)] is a seed makeup charge. The required precipitator efficiency was determined by emission requirements for all points. The seed makeup requirement was then determined by the losses in the ash as well as the stack (Appendix A 9.1). The present price of potassium carbonate [\$0.2866/kg (\$0.13/lb)] was used in calculating makeup costs. The makeup seed cost is less than 10% of the fuel costs. The overall energy cost of the base case is 7.778 mills/MJ (28 mills/kWh) with \$0.85/10^6 Btu fuel.

Table 9.13 contains a summary of the plant efficiency, capital cost, cost of electricity, and the estimated time of construction for each Base Case 2 parametric point considered.

Table 9.14 contains a summation of the major component material costs, tabulates the indirect cost, and give the variation of the cost of electricity with capacity factor, fuel costs, fixed charge, contingency and escalation for each Base Case 2 parametric point. Those major component material charges considered were: the MHD combustor (the sum of all material charges in Account 9), MHD generator duct (the sum of the materials charges in Subaccount 11.1 through 11.4), magnet and refrigerator (the sum of the materials charges in subaccounts 11.5 through 11.8), high temperature heat exchangers (all material charges in Account 13), seed recovery system (all material charges in Subaccounts 20.5, 21.1

Table 9.13 BC NO 2 OPEN CYCLE MHD-STEAM BOTTOMING SUMMARY PLANT RESULTS

PARAMETRIC POINT	1	2	3	4	5	6	7	8
THERMODYNAMIC EFF	.000	.000	.000	.000	.000	.000	.000	.000
POWER PLANT EFF	.459	.453	.457	.472	.470	.434	.465	.480
OVERALL ENERGY EFF	.469	.463	.457	.472	.470	.484	.465	.480
CAP COST MILLION $	1322.409	797.153	413.222	1236.552	1334.709	1262.609	1477.462	1347.934
CAPITAL COST,$/KWE	663.424	669.276	715.522	676.542	682.630	641.104	756.932	684.262
COE CAPITAL	23.372	21.157	22.619	20.123	21.579	20.267	23.928	21.631
COE FUEL	6.190	6.259	6.350	6.148	6.169	5.994	6.243	6.045
COE OP & MAIN	.847	.991	1.215	.801	.954	.948	1.001	.960
COST OF ELECTRIC	28.010	28.407	30.184	27.072	28.703	27.208	31.172	28.637
EST TIME OF CONST	7.000	6.397	3.644	5.998	5.941	6.977	6.967	6.980
PARAMETRIC POINT	**9**	**10**	**11**	**12**	**13**	**14**	**15**	**16**
THERMODYNAMIC EFF	.000	.000	.000	.000	.000	.000	.000	.000
POWER PLANT EFF	.455	.472	.463	.445	.455	.492	.489	.462
OVERALL ENERGY EFF	.468	.472	.468	.445	.465	.482	.489	.462
CAP COST MILLION $	1319.021	1257.910	1317.383	1352.904	1297.036	1351.887	1426.192	1304.403
CAPITAL COST,$/KWE	661.313	637.212	661.332	652.807	659.075	688.535	723.311	663.608
COE CAPITAL	20.905	20.144	20.906	21.585	20.835	21.766	22.865	20.978
COE FUEL	6.196	6.152	6.193	6.512	6.235	6.016	5.937	6.277
COE OP & MAIN	.595	.555	.461	.333	.920	.787	.820	.860
COST OF ELECTRIC	27.697	26.951	27.585	28.430	27.889	28.570	29.623	28.115
EST TIME OF CONST	7.000	6.998	7.000	7.000	6.984	6.989	6.984	6.983
PARAMETRIC POINT	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**
THERMODYNAMIC EFF	.000	.000	.000	.000	.000	.000	.000	.000
POWER PLANT EFF	.497	.000	.000	.000	.000	.000	.000	.000
OVERALL ENERGY EFF	.487	.000	.000	.000	.000	.000	.000	.000
CAP COST MILLION $	1275.593	.000	.000	.000	.000	.000	.000	.000
CAPITAL COST,$/KWE	641.657	.000	.000	.000	.000	.000	.000	.000
COE CAPITAL	20.284	.000	.000	.000	.000	.000	.000	.000
COE FUEL	5.959	.000	.000	.000	.000	.000	.000	.000
COE OP & MAIN	.900	.000	.000	.000	.000	.000	.000	.000
COST OF ELECTRIC	27.043	.000	.000	.000	.000	.000	.000	.000
EST TIME OF CONST	6.995	.000	.000	.000	.000	.000	.000	.000

Table 9.14 BC NO 2 OPEN CYCLE MHD-STEAM BOTTOMING SUMMARY PLANT RESULTS

	PARAMETRIC POINT		1	2	3	4	5	6	7	8
	TOTAL CAPITAL COST	,M$	1322.41	797.17	416.22	1266.55	1304.71	1262.61	1477.46	1347.93
P	MHD COMBUSTOR	,M$	.643	.473	.386	.543	.611	.640	.653	.641
L	MHD GENERATOR DUCT	,M$	.734	.514	.351	.569	.949	.782	1.045	1.001
A	MAGNET & REFRIGERATOR	,M$	73.151	42.734	23.953	51.341	30.241	71.851	145.629	94.968
N	HIGH TEMP HEAT EXCHANGERS	,M$	93.009	55.823	27.963	91.228	92.992	89.701	94.035	93.009
T	SEED RECOVERY SYSTEM	,M$	48.795	30.750	19.202	45.762	35.447	35.309	38.363	37.733
	INVERTER-TRANSFORMER SYSTEM	,M$	57.820	33.810	16.170	57.810	56.840	56.840	56.350	57.281
	COMPRESSOR & DRIVE	,M$	15.600	11.220	6.970	15.590	15.600	15.450	15.800	15.600
	STEAM TURB-GEN	,M$	26.580	20.260	13.015	26.780	26.580	26.090	25.880	26.080
R	TOT MAJOR COMPONENT COST	,M$	313.333	175.584	106.930	290.224	319.261	296.663	377.754	326.713
E	TOT MAJOR COMPONENT COST	,$/KWE	157.192	154.206	183.822	145.860	167.039	150.634	193.531	165.649
S	BALANCE OF PLANT COST	,$/KWE	64.035	73.434	73.071	63.981	65.221	64.353	68.019	66.207
U	SITE LABOR	,$/KWE	80.733	56.718	100.263	79.197	81.005	77.596	86.453	81.089
L	TOTAL DIRECT COST	,$/KWE	301.961	319.358	362.156	289.039	313.265	292.583	348.003	312.945
T	INDIRECT COSTS	,$/KWE	41.174	44.226	51.134	40.391	41.312	39.574	44.091	41.355
	PROF & OWNER COSTS	,$/KWE	24.157	25.549	23.372	23.123	25.061	23.407	27.840	25.036
B	CONTINGENCY COST	,$/KWE	30.196	30.011	31.306	28.897	31.141	29.192	34.685	31.233
R	ESCALATION COST	,$/KWE	119.139	113.316	111.133	114.285	121.922	114.991	135.518	122.658
E	INT DURING CONSTRUCTION	,$/KWE	140.798	136.817	130.821	140.808	149.929	141.456	166.795	151.035
A	TOTAL CAPITALIZATION	,$/KWE	663.424	659.275	715.522	636.542	682.533	641.104	756.932	684.262
K	COST OF ELEC-CAPITAL	,MILLS/KWE	20.972	21.157	22.619	20.123	21.579	20.267	23.928	21.631
D	COST OF ELEC-FUEL	,MILLS/KWE	6.199	6.259	6.350	6.148	6.169	5.994	6.243	6.045
O	COST OF ELEC-OP&MAIN	,MILLS/KWE	.847	.991	1.215	.801	.954	.948	1.001	.960
W	TOTAL COST OF ELEC	,MILLS/KWE	28.018	28.407	30.184	27.072	28.703	27.208	31.172	28.637
N	COE 0.5 CAP. FACTOR	,MILLS/KWE	34.302	34.754	36.970	33.108	35.176	33.288	38.350	35.126
	COE 0.8 CAP. FACTOR	,MILLS/KWE	24.073	24.440	25.943	23.299	24.656	23.408	26.685	24.591
	COE 1.2XCAP. COST	,MILLS/KWE	32.204	32.639	34.708	31.096	33.018	31.262	35.958	32.963
	COE 1.2XFUEL COST	,MILLS/KWE	29.249	29.659	31.454	28.301	29.936	28.407	32.420	29.846
	COE (CONTINGENCY=0)	,MILLS/KWE	26.417	26.892	28.689	25.547	27.067	25.671	29.346	26.991
	COE (ESCALATION=0)	,MILLS/KWE	23.332	24.041	25.001	22.595	23.924	22.700	25.856	23.824

	PARAMETRIC POINT		9	10	11	12	13	14	15	16
	TOTAL CAPITAL COST	,M$	1318.02	1167.91	1317.98	1352.90	1297.04	1361.89	1426.19	1304.40
P	MHD COMBUSTOR	,M$	.649	.653	.825	.644	.606	.529	.457	.643
L	MHD GENERATOR DUCT	,M$	.734	.567	.734	.734	.717	.981	1.026	.734
A	MAGNET & REFRIGERATOR	,M$	70.151	51.841	73.151	70.151	79.151	101.922	140.762	70.151
N	HIGH TEMP HEAT EXCHANGERS	,M$	93.009	71.643	93.009	93.009	89.316	87.310	79.351	93.009
T	SEED RECOVERY SYSTEM	,M$	46.370	45.495	46.060	58.641	46.572	45.147	44.559	48.796
	INVERTER-TRANSFORMER SYSTEM	,M$	57.820	57.820	57.820	57.820	57.820	63.210	66.640	57.820
	COMPRESSOR & DRIVE	,M$	15.600	15.590	15.500	15.600	15.700	16.220	17.200	15.600
	STEAM TURB-GEN	,M$	26.580	26.780	26.580	26.580	25.480	24.570	24.572	26.580
R	TOT MAJOR COMPONENT COST	,M$	310.912	290.390	310.779	323.179	306.361	339.889	374.567	313.333
E	TOT MAJOR COMPONENT COST	,$/KWE	155.999	145.941	155.941	163.107	155.674	171.839	189.966	159.406
S	BALANCE OF PLANT COST	,$/KWE	64.113	64.047	64.148	63.792	64.378	62.765	61.845	62.507
U	SITE LABOR	,$/KWE	80.763	79.330	80.786	83.645	80.314	80.515	80.937	80.637
L	TOTAL DIRECT COST	,$/KWE	300.875	289.317	300.875	310.544	300.367	315.119	332.749	302.551
T	INDIRECT COSTS	,$/KWE	41.189	40.458	41.201	42.659	40.960	41.063	41.278	41.125
	PROF & OWNER COSTS	,$/KWE	24.073	23.145	24.070	24.944	24.029	25.210	26.620	24.204
B	CONTINGENCY COST	,$/KWE	30.088	20.925	30.088	31.054	29.989	31.477	33.223	30.205
R	ESCALATION COST	,$/KWE	116.759	114.407	118.762	122.619	118.184	123.523	129.706	118.990
E	INT DURING CONSTRUCTION	,$/KWE	146.331	140.958	146.335	151.087	145.545	152.144	159.735	146.534
A	TOTAL CAPITALIZATION	,$/KWE	651.313	637.212	561.332	682.807	559.075	688.535	723.311	663.608
K	COST OF ELEC-CAPITAL	,MILLS/KWE	20.905	20.144	20.906	21.585	20.835	21.766	22.865	20.978
D	COST OF ELEC-FUEL	,MILLS/KWE	6.195	5.152	5.198	6.512	6.235	6.016	5.937	6.277
O	COST OF ELEC-OP&MAIN	,MILLS/KWE	.596	.656	.481	.333	.820	.787	.820	.860
W	TOTAL COST OF ELEC	,MILLS/KWE	27.697	25.951	27.585	28.430	27.889	28.570	29.523	28.115
N	COE 0.5 CAP. FACTOR	,MILLS/KWE	33.969	32.394	33.857	34.905	34.139	35.099	36.483	34.409
	COE 0.8 CAP. FACTOR	,MILLS/KWE	23.777	23.174	23.665	24.393	23.982	24.488	25.336	24.182
	COE 1.2XCAP. COST	,MILLS/KWE	31.878	30.960	31.767	32.747	32.056	32.923	34.196	32.311
	COE 1.2XFUEL COST	,MILLS/KWE	29.936	28.192	28.825	29.732	29.136	29.773	30.811	29.371
	COE (CONTINGENCY=0)	,MILLS/KWE	26.110	25.425	25.998	26.791	26.309	26.910	27.872	26.523
	COE (ESCALATION=0)	,MILLS/KWE	23.034	22.450	22.922	23.615	23.251	23.721	24.533	23.446

Table 9.14 ECAS NO 2 OPEN CYCLE MHD-STEAM BOTTOMING SUMMARY PLANT RESULTS
Continued

	PARAMETRIC POINT		17	18	19	20	21	22	23	24
	TOTAL CAPITAL COST	,M$	1275.59	.00	.00	.00	.00	.00	.00	.00
PLANT	MHD COMBUSTOR	,M$	.631	.000	.000	.000	.000	.000	.000	.000
	MHD GENERATOR DUCT	,M$	.599	.000	.000	.000	.000	.000	.000	.000
	MAGNET & REFRIGERATOR	,M$	55.815	.000	.000	.000	.000	.000	.000	.000
	HIGH TEMP HEAT EXCHANGERS	,M$	88.043	.000	.000	.000	.000	.000	.000	.000
	SEED RECOVERY SYSTEM	,M$	44.693	.000	.000	.000	.000	.000	.000	.000
	INVERTER-TRANSFORMER SYSTEM	,M$	60.270	.000	.000	.000	.000	.000	.000	.000
	COMPRESSOR & DRIVE	,M$	15.880	.000	.000	.000	.000	.000	.000	.000
	STEAM TURB-GEN	,M$	24.862	.000	.000	.000	.000	.000	.000	.000
RESULT BREAKDOWN	TOT MAJOR COMPONENT COST	,M$	330.791	.000	.000	.000	.000	.000	.000	.000
	TOT MAJOR COMPONENT COST	,$/KWE	151.306	.000	.000	.000	.000	.000	.000	.000
	BALANCE OF PLANT COST	,$/KWE	62.790	.000	.000	.000	.000	.000	.000	.000
	SITE LABOR	,$/KWE	78.143	.000	.000	.000	.000	.000	.000	.000
	TOTAL DIRECT COST	,$/KWE	292.149	.000	.000	.000	.000	.000	.000	.000
	INDIRECT COSTS	,$/KWE	39.853	.000	.000	.000	.000	.000	.000	.000
	PROF & OWNER COSTS	,$/KWE	23.372	.000	.000	.000	.000	.000	.000	.000
	CONTINGENCY COST	,$/KWE	29.202	.000	.000	.000	.000	.000	.000	.000
	ESCALATION COST	,$/KWE	115.191	.000	.000	.000	.000	.000	.000	.000
	INT DURING CONSTRUCTION	,$/KWE	141.900	.000	.000	.000	.000	.000	.000	.000
	TOTAL CAPITALIZATION	,$/KWE	541.657	.000	.000	.000	.000	.000	.000	.000
	COST OF ELEC-CAPITAL	,MILLS/KWE	20.284	.000	.000	.000	.000	.000	.000	.000
	COST OF ELEC-FUEL	,MILLS/KWE	5.959	.000	.000	.000	.000	.000	.000	.000
	COST OF ELEC-OP&MAIN	,MILLS/KWE	.800	.000	.000	.000	.000	.000	.000	.000
	TOTAL COST OF ELEC	,MILLS/KWE	27.043	.000	.000	.000	.000	.000	.000	.000
	COE 0.5 CAP. FACTOR	,MILLS/KWE	33.123	.000	.000	.000	.000	.000	.000	.000
	COE 0.8 CAP. FACTOR	,MILLS/KWE	23.233	.000	.000	.000	.000	.000	.000	.000
	COE 1.2XCAP. COST	,MILLS/KWE	31.099	.000	.000	.000	.000	.000	.000	.000
	COE 1.2XFUEL COST	,MILLS/KWE	29.234	.000	.000	.000	.000	.000	.000	.000
	COE (CONTINGENCY=0)	,MILLS/KWE	25.502	.000	.000	.000	.000	.000	.000	.000
	COE (ESCALATION=0)	,MILLS/KWE	22.521	.000	.000	.000	.000	.000	.000	.000

and 21.3), dc inverter-transformer subsystem (the sum of the material charges in Subaccounts 15.1 and 15.2), compressor and steam turbine drive (material charges in subaccounts 11.9 and 11.10) and steam turbine generator (material charges in Subaccounts 11.11 and 11.12).

Table 9.15 is the natural resource summary table for each Base Case 2 parametric point.

Points 2 and 3 are simply scaled down versions of the base case. The power output of Point 2 is 1191 MW, about 60% of the base case. The specific capital cost ($/kW) is about 1% greater than the base case, and the efficiency is about 0.6 of a point lower. The energy cost is 7.889 mills/MJ (28.4 mills/kWh). Point 3 has a power output of 582 MW or about 30% of the base case. The specific capital cost is about 8% higher, and the efficiency nearly a point lower than the base case. Also, the O&M is nearly 0.4 mill higher due to the increased cost of manning the smaller plant. The energy cost is about 0.611 mills/MJ (2.2 mills/kWh) higher than for the base case.

Point 4 uses the bituminous coal dried to a 3% moisture content. The removal of moisture results in a higher flame temperature (for a given air preheat temperature) and also produces a better plasma electrical conductivity for a given temperature. As a result, the efficiency is about 0.3 point higher, the capital charge about 0.8 mill lower, and O&M about 0.0111 mill/MJ (0.04 mill/kWh) lower. The energy cost is reduced to about 7.5 mills/MJ (27 mills/kWh).

For Points 5 and 6, the fuels are the subbituminous coal with 20% and 16% moisture, respectively. The higher moisture content reduces the combustor temperature and conductivity, increasing the magnet cost and decreasing the basic plant efficiency. The lower sulfur content, however, reduces the equipment, energy, and power requirements of the seed treatment plant so that the overall efficiency and total energy costs are competitive with the base case. The lower moisture content of the Point 6 fuel gives it a 0.4167 mill/MJ (1.5 mills/kWh) advantage over Point 5.

Table 9.15 BC NO 2 OPEN CYCLE MHD-STEAM BOTTOMING NATURAL RESOURCE REQUIREMENTS

PARAMETRIC POINT	1	2	3	4	5	6	7	8
COAL, LB/KW-HR	.67506	.68258	.69254	.67047	.81151	.78841	1.06594	1.03227
SORBANT OR SEED,LB/KW-HR	.00420	.00427	.00372	.00385	.00498	.00498	.00535	.00506
TOTAL WATER, GAL/KW-HR	.582	.597	.612	.580	.558	.568	.576	.569
COOLING WATER	.567	.592	.596	.565	.550	.560	.567	.560
GASIFIER PROCESS H2O	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
CONDENSATE MAKE UP ,	.01111	.01127	.01179	.01101	.00693	.00711	.00765	.00750
WASTE HANDLING SLURRY	.0040	.0040	.0043	.0039	.0011	.0011	.0015	.0014
SCRUBBER WASTE WATER	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
NOX SUPPRESSION	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
TOTAL LAND ACRES/100MWE	51.30	56.08	65.99	51.25	48.33	47.98	49.64	48.95
MAIN PLANT	11.64	15.86	24.38	11.65	11.90	11.69	11.76	11.70
DISPOSAL LAND	14.12	14.27	14.48	14.02	11.69	11.36	12.73	12.33
LAND FOR ACCESS RR	25.54	25.95	28.13	25.59	24.73	24.93	25.15	24.92

PARAMETRIC POINT	9	10	11	12	13	14	15	16
COAL, LB/KW-HR	.67568	.67086	.67595	.71016	.67990	.65610	.64749	.68456
SORBANT OR SEED,LB/KW-HR	.00230	.00276	.00144	.00027	.00398	.00376	.00403	.00426
TOTAL WATER, GAL/KW-HR	.583	.580	.583	.598	.578	.546	.530	.600
COOLING WATER	.567	.565	.567	.574	.564	.532	.516	.585
GASIFIER PROCESS H2O	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
CONDENSATE MAKE UP ,	.01116	.01110	.01120	.01597	.01060	.01058	.01033	.01113
WASTE HANDLING SLURRY	.0040	.0039	.0040	.0075	.0036	.0038	.0038	.0040
SCRUBBER WASTE WATER	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
NOX SUPPRESSION	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
TOTAL LAND ACRES/100MWE	51.31	51.26	51.32	52.15	50.89	49.31	48.30	51.95
MAIN PLANT	11.64	11.65	11.64	11.71	11.73	11.69	11.71	11.74
DISPOSAL LAND	14.13	14.03	14.13	14.75	14.22	13.72	13.54	14.32
LAND FOR ACCESS RR	25.54	25.59	25.54	25.69	24.95	23.90	23.05	25.90

PARAMETRIC POINT	17	18	19	20	21	22	23	24
COAL, LB/KW-HR	.64991	.00000	.00000	.00000	.00000	.00000	.00000	.00000
SORBANT OR SEED,LB/KW-HR	.00395	.00000	.00000	.00000	.00000	.00000	.00000	.00000
TOTAL WATER, GAL/KW-HR	.564	.000	.000	.000	.000	.000	.000	.000
COOLING WATER	.550	.000	.000	.000	.000	.000	.000	.000
GASIFIER PROCESS H2O	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
CONDENSATE MAKE UP ,	.01067	.00000	.00000	.00000	.00000	.00000	.00000	.00000
WASTE HANDLING SLURRY	.0038	.0000	.0000	.0000	.0000	.0000	.0000	.0000
SCRUBBER WASTE WATER	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
NOX SUPPRESSION	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
TOTAL LAND ACRES/100MWE	49.94	.00	.00	.00	.00	.00	.00	.00
MAIN PLANT	11.65	.00	.00	.00	.00	.00	.00	.00
DISPOSAL LAND	13.59	.00	.00	.00	.00	.00	.00	.00
LAND FOR ACCESS RR	24.69	.00	.00	.00	.00	.00	.00	.00

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The fuels for Points 7 and 8 are the lignite coals with 27% and 18% moisture respectively. The combustor temperature is down significantly, and the cost of the magnet becomes rather large and cannot be offset by the savings in seed treatment. The energy cost of both points is higher than that for the base case. A similar problem would face any type of power plant which burns an inferior fuel, but it is a more severe problem in the open-cycle MHD plant since the products of combustion are the generator working fluid. The situation can be improved by increased drying of the coal and/or higher air preheat temperatures. A lower pressure ratio might also be desirable for these lower temperatures.

Point 9 demonstrates the effect of a decreased ash carry-over from the combustor. Since less seed is combined with the ash, the seed makeup costs are reduced by about 0.25 mill. Since there is less fly ash to collect, the precipitator cost is reduced significantly. Since more seed must be treated, however, more power and energy are required for seed treatment, and the efficiency of the plant is reduced very slightly from the base case. The overall energy cost is about 0.0833 mill/MJ (0.3 mill/kWh) lower than for the base case.

Point 10 combines the dried bituminous coal (3% moisture) with reduced ash carry-over. The energy costs are a 0.278 mill/MJ (1 mill/kWh) lower than for the base case, for the reasons discussed under Points 4 and 9.

For Point 11, the ash carry-over is reduced further to 5% with the bituminous coal (13% moisture). Due to a further reduction in seed makeup (below that of Point 9), the energy cost is reduced to 7.667 mills/MJ (27.6 mills/kWh).

For Point 12, all of the ash is carried over to the MHD generator. This case requires a leaching plant to separate the potassium from the ash (Appendix A 9.1). The equipment, energy, and power requirements for seed treatment are increased, but this is largely offset by the reduced makeup required. The makeup is reduced because nearly all of the potassium can be leached from the ash. For points with reduced ash carry-over, it was felt that the seed loss was too small to justify a leaching operation. The total energy costs are 0.11 mill/MJ (0.4 mill/kWh) higher than for the base case.

In Point 13, some exhaust products are recycled with the air stream to reduce the combustor temperature to 2700°K (4400°F). There is a slight decrease in efficiency due to the reduced temperature and a slight reduction in capital cost due to the larger portion of the power generated by the steam plant. The energy cost is about 0.0278 mill/MJ (0.1 mill/kWh) lower than the base case.

In Points 14 and 15 the combustor pressures are raised to 0.8106 and 1.0133 MPa (8 and 10 atm) respectively. The increased magnet cost raises the capital cost per kilowatt about 4% and 9% above the base case respectively. This increased cost more than offsets the fuel savings for the standard conditions, and the total energy costs are 0.1667 and 0.4444 mill/MJ (0.6 and 1.6 mills/kWh) higher than the base case for Points 14 and 15 respectively.

In Point 16, a 16.6 MPa (2400 psi) steam plant is substituted for the 24.1 MPa (3500 psi) plant of the base case. There is a reduction in total plant cost, but this is more than offset by a reduction in power output, and the specific plant cost increases slightly. There is also a decrease in plant efficiency and the total energy cost is 0.0278 mill/MJ (0.1 mill/kWh) higher than the base case.

Point 17 uses the dried bituminous coal with an MHD pressure ratio of seven. Compared to Point 4, the efficiency is about 1-1/2 points better. The specific capital cost, however, increases by nearly 1%, and the resultant energy costs are only 0.0083 mill/MJ (0.03 mill/kWh) better with the standard set of economic parameters.

9.4.2 Base Case 3 and Variations

The detailed accounts listing, cost of electricity summary, and input-output sheets for Base Case 3 are given as Tables 9.16, 9.17, and 9.18, respectively. In this case, 76.44% of the total direct cost is found in Accounts 8, 11, 12, 13, and 15 with 20% of the direct cost attributable to the gasification subsystem. The total direct costs of some of the major component groupings were $163,000,000 for the gasifier, $182,500,000 for the superconducting magnet; $18,960,000 for the steam turbine-driven compressor; $25,400,000 for the steam turbine generator; $54,920,000 for the heat recovery steam generator; $76,240,000 for the air heater system; and $73,840,000 for the inverter filter system.

Table 9.16 3C NO 3 OPEN CYCLE MHD-STEAM BOTTOMING ACCOUNT LISTING
PARAMETRIC POINT NO. 1

ACCOUNT NO. & NAME,	UNIT	AMOUNT	MAT $/UNIT	INS $/UNIT	MAT COST,$	INS COST,$
SITE DEVELOPMENT						
1. 1 LAND COST	ACRE	234.0	1000.00	.00	234000.00	.00
1. 2 CLEARING LAND	ACRE	73.0	.00	600.00	.00	45795.32
1. 3 GRADING LAND	ACRE	234.0	.00	3000.00	.00	702000.00
1. 4 ACCESS RAILROAD	MILE	5.0	115000.00	110000.00	575000.00	550000.00
1. 5 LOOP RAILROAD TRACK	MILE	3.0	120000.00	70000.00	360000.00	210000.00
1. 6 SIDING R R TRACK	MILE	.0	125000.00	80000.00	.00	.00
1. 7 OTHER SITE COSTS	ACRE	.0	.00	.00	482693.86	482693.86
PERCENT TOTAL DIRECT COST IN ACCOUNT 1 = .463				ACCOUNT TOTAL,$	1651693.86	1991489.17
EXCAVATION & PILING						
2. 1 COMMON EXCAVATION	YD3	205500.0	.00	3.00	.00	616500.00
2. 2 PILING	FT	548000.0	6.50	8.50	3562000.00	4658000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 2 = 1.123				ACCOUNT TOTAL,$	3562000.00	5274500.00
PLANT ISLAND CONCRETE						
3. 1 PLANT IS. CONCRETE	YD3	68500.0	70.00	80.00	4795000.00	5480000.00
3. 2 SPECIAL STRUCTURES	YD3	.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 3 = 1.306				ACCOUNT TOTAL,$	4795000.00	5480000.00
HEAT REJECTION SYSTEM						
4. 1 COOLING TOWERS	EACH	24.0	.00	.00	3684000.00	1836000.00
4. 2 CIRCULATING H2O SYS	EACH	1.0	.00	.00	1600402.98	2145938.12
4. 3 SURFACE CONDENSER	FT2	583987.7	.00	.00	2739904.91	408791.40
PERCENT TOTAL DIRECT COST IN ACCOUNT 4 = 1.578				ACCOUNT TOTAL,$	8024307.81	4390729.50
STRUCTURAL FEATURES						
5. 1 STAT. STRUCTURAL ST.	TON	3550.0	650.00	175.00	2307500.00	621250.00
5. 2 SILOS & BUNKERS	TPH	.0	1800.00	750.00	.00	.00
5. 3 CHIMNEY	FT	400.0	.00	.00	435070.92	552606.38
5. 4 STRUCTURAL FEATURES	EACH	1.0	862000.00	157000.00	862000.00	157000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 5 = .640				ACCOUNT TOTAL,$	3604570.91	1330856.37
BUILDINGS						
6. 1 STATION BUILDINGS	FT3	9312000.0	.16	.16	1489920.00	1489920.00
6. 2 ADMINSTRATION	FT2	15000.0	16.00	14.00	240000.00	210000.00
6. 3 WAREHOUSE & SHOP	FT2	24000.0	12.00	8.00	288000.00	192000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 6 = .497				ACCOUNT TOTAL,$	2017920.00	1891920.00
FUEL HANDLING & STORAGE						
7. 1 COAL HANDLING SYS	TPH	599.4	.00	.00	11209788.37	4586668.62
7. 2 DOLOMITE HAND. SYS	TPH	316.6	.00	.00	4536129.37	2003197.80
7. 3 FUEL OIL HAND. SYS	GAL	2250000.0	.00	.00	258657.35	202919.16
PERCENT TOTAL DIRECT COST IN ACCOUNT 7 = 2.897				ACCOUNT TOTAL,$	16004575.00	6792785.50
FUEL PROCESSING						
8. 1 COAL DRYER & CRUSHER	TPH	.0	.00	.00	.00	.00
8. 2 CARBONIZERS	TPH	.0	.00	.00	.00	.00
8. 3 GASIFIERS	TPH	599.4	.00	.00	104358864.00	58701861.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 8 = 20.724				ACCOUNT TOTAL,$	104358864.00	58701861.00

Table 9.16 Continued

BC NO 3 OPEN CYCLE MHD-STEAM BOTTOMING ACCOUNT LISTING

PARAMETRIC POINT NO. 1

ACCOUNT NO. & NAME,	UNIT	AMOUNT	MAT $/UNIT	INS $/UNIT	MAT COST,$	INS COST,$
FIRING SYSTEM						
9. 1 CONTAINMENT STEEL	KP	268.0	350.00	159.00	93800.00	42612.00
9. 2 BRICK- DENSE ZRO2	KP	144.0	3305.00	150.00	475920.00	21600.00
9. 3 BRICK-SILICON CARBIDE	KP	16.4	1400.00	150.00	22960.00	2460.00
9. 4 BRICK- MGO	KP	96.0	131.00	150.00	12576.00	14400.00
9. 5 BRICK (INSULATING)	KP	18.5	159.00	150.00	2941.50	2775.00
9. 6 STRUCTURAL STEEL	KP	163.0	509.00	150.00	82967.00	24450.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 9 = .102 ACCOUNT TOTAL,$					691164.50	108297.00
VAPOR GENERATOR (FIRED)						
10. 1		.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 10 = .000 ACCOUNT TOTAL,$					.00	.00
ENERGY CONVERTER						
11. 1 DUCT INSULATION BRICK	KP	49.7	137.00	274.00	6808.90	13617.80
11. 2 DUCT CONDUCTING BRICK	KP	49.7	500.00	1225.00	24850.00	60882.50
11. 3 DUCT STRUCTURAL STEEL	KP	283.0	3000.00	3000.00	849000.00	849000.00
11. 4 DUCT COOLING TUBES	KP	292.0	3000.00	3000.00	876000.00	876000.00
11. 5 SUPERCONDUCTING MAT	EACH	1.0	77599999.00	19399999.75	77599999.00	19399999.75
11. 6 MAGNET STRUCT STEEL	EACH	1.0	85499999.00	.00	85499999.00	.00
11. 7 NITROGEN LIQUIFIER	EACH	1.0	207000.00	23000.00	207000.00	23000.00
11. 8 HELIUM LIQUIFIER	EACH	1.0	701999.99	78000.00	701999.99	78000.00
11. 9 COMPRESSOR	EACH	1.0	5590000.00	558999.99	5590000.00	558999.99
11.10 COMPRESSOR DRIVE	EACH	1.0	11900000.00	908396.26	11900000.00	908396.26
11.11 STEAM TURB-GEN	EACH	1.0	23800000.00	1592734.39	23800000.00	1592734.39
11.12 FEED WATER HEATERS	PLANT	1.0	530000.00	15900.00	530000.00	15900.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 11 =29.481 ACCOUNT TOTAL,$					207585652.00	24376529.75
COUPLING HEAT EXCHANGER						
12. 1 HIGH TEMP SUPERHEATER	EA	1.0	3941999.97	1970999.98	3941999.97	1970999.98
12. 2 REHEATER & EVAP SECT	EA	1.0	8640000.00	5759999.94	8640000.00	5759999.94
12. 3 ECONOMIZER SECTION	EA	1.0	24222240.00	10380960.00	24222240.00	10380960.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 12 = 6.979 ACCOUNT TOTAL,$					36804239.50	18111959.75

Table 9.16   BC NO 3 OPEN CYCLE MHD-STEAM BOTTOMING ACCOUNT LISTING
Continued   PARAMETRIC POINT NO. 1

ACCOUNT NO. & NAME,	UNIT	AMOUNT	MAT $/UNIT	INS $/UNIT	MAT COST,$	INS COST,$
HEAT RECOVERY HEAT EXCH.						
13. 1 CERAMIC TUBING	KP	725.0	8640.00	1020.00	6264000.00	739500.00
13. 2 EXOTIC METAL TUBES	KP	2920.0	11316.00	6720.00	33042720.00	19622400.00
13. 3 STAINLESS STEEL TUBES	KP	2699.0	2940.00	1800.00	7935060.00	4858200.00
13. 4 TUBE CERAMIC COATINGS	KP	168.0	585.60	513.60	98380.80	86284.80
13. 5 INSULATION FOR REGEN	KP	2397.0	230.40	180.00	552268.80	431460.00
13. 6 STRUCTURAL STEEL	KP	111.0	564.00	187.20	62604.00	20779.20
13. 7 CONT STEEL REGEN	KP	1137.0	420.00	190.80	477540.00	216939.60
13. 8 CHECKER BRICKS	KP	.0	.00	.00	.00	.00
13. 9 INSULATION SFA HEATER	KP	.0	.00	.00	.00	.00
13.10 CONT ST SFA HEATER	KP	.0	.00	.00	.00	.00
13.11 STRU ST SFA HEATER	KP	.0	.00	.00	.00	.00
13.12 BURNER SFA HEATER	EACH	.0	.00	.00	.00	.00
13.13 HIGH TEMP VALVES	EACH	.0	.00	.00	.00	.00
13.14 PREHEATER AIR PIPING	TON	.0	.00	.00	.00	.00
13.15 FUEL GAS PIPING	TON	.0	.00	.00	.00	.00
13.16 COMB AIR PIPING	TON	.0	.00	.00	.00	.00
13.17 RECIRC AIR PIPING	TON	.0	.00	.00	.00	.00
13.18 PRODUCTS PIPING	TON	.0	.00	.00	.00	.00
13.19 RECIRC ASPIRATOR	TON	.0	.00	.00	.00	.00
13.20 HEADERS	KP	450.0	912.00	2136.00	410400.00	961200.00
13.21 SEALS	EACH	5189.0	20.40	40.80	105855.60	211711.20
13.22 CONCRETE	YD3	116.0	84.00	96.00	9744.00	11136.00
13.23 EXPANSION JOINTS	EACH	2.0	48000.00	12000.00	96000.00	24000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 13 = 3.639			ACCOUNT	TOTAL,$	43054572.50	27183610.25
WATER TREATMENT						
14. 1 DEMINERALIZER	GPH	1329.9	2000.00	560.00	2659803.22	744744.90
14. 2 CONDENSATE POLISHING	KWE	358700.0	1.25	.30	1210875.00	290610.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 14 = .624			ACCOUNT	TOTAL,$	3870678.22	1035354.90
POWER CONDITIONING						
15. 1 INVERTER SYSTEM	KWE	1356700.0	43.00	4.77	58339099.00	6478242.37
15. 2 FILTERS	KWE	1356700.0	6.00	.65	8140199.87	881854.98
15. 3 STD TRANSFORMERS	KWE	733700.0	.00	.00	1416436.97	28328.74
PERCENT TOTAL DIRECT COST IN ACCOUNT 15 = 9.568			ACCOUNT	TOTAL,$	67894735.00	7388426.00
AUXILIARY MECH EQUIPMENT						
16. 1 BOILER FEED PUMP &DR.	KWE	920265.0	1.67	.10	1536842.55	92026.50
16. 2 OTHER PUMPS	KWE	743660.0	.88	.12	654420.79	89239.20
16. 3 MISC SERVICE SYS	KWE	1859150.0	1.17	.73	2175205.50	1357179.48
16. 4 AUXILIARY BOILER	PPH	720000.0	4.00	.80	2880000.00	575999.99
PERCENT TOTAL DIRECT COST IN ACCOUNT 16 = 1.190			ACCOUNT	TOTAL,$	7246468.81	2114445.16
PIPE & FITTINGS						
17. 1 CONVENTIONAL PIPING	TON	3635.0	3000.00	1800.00	10905000.00	6543000.00
17. 2 HIGT TEMP AIR PIPING	TON	3500.0	450.00	225.00	1575000.00	787500.00
17. 3 LOW TEMP AIR PIPING	TON	110.0	1200.00	800.00	132000.00	88000.00
17. 4 RECIR PRODUCT PIPING	TON	.0	1200.00	1100.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 17 = 2.546			ACCOUNT	TOTAL,$	12612000.00	7418500.00

Table 9.16 Continued

3C NO 3 OPEN CYCLE MHD-STEAM BOTTOMING ACCOUNT LISTING
PARAMETRIC POINT NO. 1

ACCOUNT NO. & NAME,	UNIT	AMOUNT	MAT $/UNIT	INS $/UNIT	MAT COST,$	INS COST,$
AUXILIARY ELEC EQUIPMENT						
18. 1 MISC MOTERS,ETC		1301405.0	1.40	.17	1821967.00	221238.85
18. 2 SWITCHGEAR & MCC PAN	KWE	1301405.0	1.95	.45	2537739.75	585632.24
18. 3 CONDUIT,CABLES,TRAYS	FT	5100000.0	1.32	1.36	6731999.94	6935999.94
18. 4 ISOLATED PHASE BUS	FT	570.0	510.00	450.00	290700.00	256500.00
18. 5 LIGHTING & COMMUN.	KWE	1859150.0	.35	.43	650702.50	799434.49
PERCENT TOTAL DIRECT COST IN ACCOUNT 18 = 2.548 ACCOUNT TOTAL,$					12033109.12	8798805.37
CONTROL, INSTRUMENTATION						
19. 1 COMPUTER	EACH	1.0	726000.00	16500.00	726000.00	16500.00
19. 2 OTHER CONTROLS	EACH	1.0	1390000.00	835000.00	1390000.00	835000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 19 = .377 ACCOUNT TOTAL,$					2116000.00	851500.00
PROCESS WASTE SYSTEMS						
20. 1 BOTTOM ASH	TPH	58.7	5512793.37	1378195.84	5512793.37	1378195.84
20. 2 DRY ASH	TPH	.0	.00	.00	.00	.00
20. 3 WET SLURRY	TPH	316.6	8036551.37	2009137.84	8036551.37	2009137.84
20. 4 ONSITE DISPOSAL	ACRE	1053.7	5058.11	7714.48	5329922.37	8129031.94
20. 5 SEED TREATMENT	EACH	1.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 20 = 3.863 ACCOUNT TOTAL,$					18879257.00	11516365.62
STACK GAS CLEANING						
21. 1 PRECIPITATOR	EACH	1.0	21800000.00	3360000.00	21800000.00	3360000.00
21. 2 SCRUBBER	KWE	.0	14.87	6.41	.00	.00
21. 3 MISC STEEL & DUCTS		1.0	4000000.00	.00	4000000.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 21 = 3.706 ACCOUNT TOTAL,$					25800000.00	3360000.00
TOTAL DIRECT COSTS,$					588606792.00	198217928.00

Table 9.17
BC NO 3 OPEN CYCLE MHD-STEAM BOTTOMING COST OF ELECTRICITY,MILLS/KW.HR
PARAMETRIC POINT NO. 1

ACCOUNT	RATE, PERCENT	LABOR RATE, $/HR 6.00	8.50	10.60	15.00	21.50
TOTAL DIRECT COSTS,$	.0	703805616.	747555129.	786824712.	869103856.	990052584.
INDIRECT COST,$	51.0	57221401.	81063651.	101091141.	143053502.	205043352.
PROF & OWNER COSTS,$	8.0	56054449.	59804410.	62945977.	69528308.	79252206.
CONTINGENCY COST,$	10.0	70080561.	74755512.	78682470.	86910385.	99065258.
SUB TOTAL,$	.0	834172324.	953178695.	1029544288.	1168596032.	1374013376.
ESCALATION COST,$	6.5	265010708.	238691184.	308582780.	350260416.	411829652.
INTREST DURING CONST,$	10.0	326538056.	355716412.	380226224.	431580132.	507443852.
TOTAL CAPITALIZATION,$	.0	1475720768.	1607586272.	1718353280.	1950436576.	2293286848.
COST OF ELEC-CAPITAL	18.0	24.74022	26.95092	28.80791	32.69875	38.44658
COST OF ELEC-FUEL	.0	5.98449	5.98449	5.98449	5.98449	5.98449
COST OF ELEC-OP & MAIN	.0	3.51823	3.51823	3.51823	3.51823	3.51823
TOTAL COST OF ELEC	.0	34.24293	36.45363	38.31063	42.20146	47.94929

ACCOUNT	RATE, PERCENT	CONTINGENCY, PERCENT -5.00	.00	10.00	5.00	20.00
TOTAL DIRECT COSTS,$	.0	786824712.	786824712.	786824712.	786824712.	786824712.
INDIRECT COST,$	51.0	101091141.	101091141.	101091141.	101091141.	101091141.
PROF & OWNER COSTS,$	8.0	62945977.	62945977.	62945977.	62945977.	62945977.
CONTINGENCY COST,$	20.0	-39341235.	0.	78682470.	39341235.	157364940.
SUB TOTAL,$	.0	911520592.	950861824.	1029544288.	990203056.	1108226752.
ESCALATION COST,$	5.5	273207828.	294999475.	308582780.	296791128.	332166080.
INTREST DURING CONST,$	10.0	336638296.	351167604.	380226224.	365696916.	409284844.
TOTAL CAPITALIZATION,$	.0	1521366704.	1572029895.	1718353280.	1652691088.	1849677664.
COST OF ELEC-CAPITAL	18.0	25.50547	26.60628	28.80791	27.70710	31.00955
COST OF ELEC-FUEL	.0	5.98449	5.98449	5.98449	5.98449	5.98449
COST OF ELEC-OP & MAIN	.0	3.51823	3.51823	3.51823	3.51823	3.51823
TOTAL COST OF ELEC	.0	35.00818	35.10899	38.31063	37.20981	40.51226

ACCOUNT	RATE, PERCENT	ESCALATION RATE, PERCENT 5.00	6.50	8.00	10.00	.00
TOTAL DIRECT COSTS,$	.0	786824712.	786824712.	786824712.	786824712.	786824712.
INDIRECT COST,$	51.0	101091141.	101091141.	101091141.	101091141.	101091141.
PROF & OWNER COSTS,$	8.0	62945977.	62945977.	62945977.	62945977.	62945977.
CONTINGENCY COST,$	10.0	78682470.	78682470.	78682470.	78682470.	78682470.
SUB TOTAL,$	.0	1029544288.	1029544288.	1029544288.	1029544288.	1029544288.
ESCALATION COST,$	.0	231042160.	308582780.	390219224.	505785280.	0.
INTREST DURING CONST,$	10.0	361742516.	380226224.	399504572.	426492208.	305535272.
TOTAL CAPITALIZATION,$	.0	1622329056.	1718353280.	1819268064.	1961781776.	1335079552.
COST OF ELEC-CAPITAL	18.0	27.19808	28.80791	30.49973	32.88895	22.38239
COST OF ELEC-FUEL	.0	5.98449	5.98449	5.98449	5.98449	5.98449
COST OF ELEC-OP & MAIN	.0	3.51823	3.51823	3.51823	3.51823	3.51823
TOTAL COST OF ELEC	.0	36.70079	38.31063	40.00245	42.39166	31.88510

ACCOUNT	RATE, PERCENT	INT DURING CONST,PERCENT 6.00	8.00	10.00	12.50	15.00
TOTAL DIRECT COSTS,$	.0	786824712.	786824712.	786824712.	786824712.	786824712.
INDIRECT COST,$	51.0	101091141.	101091141.	101091141.	101091141.	101091141.
PROF & OWNER COSTS,$	8.0	62945977.	62945977.	62945977.	62945977.	62945977.
CONTINGENCY COST,$	10.0	78682470.	78682470.	78682470.	78682470.	78682470.
SUB TOTAL,$	.0	1029544288.	1029544288.	1029544288.	1029544288.	1029544288.
ESCALATION COST,$	6.5	308582780.	308582780.	308582780.	308582780.	308582780.
INTREST DURING CONST,$	15.0	218098266.	297403236.	380226224.	486909604.	603577200.
TOTAL CAPITALIZATION,$	.0	1555225312.	1635530288.	1718353280.	1827036656.	1941704256.
COST OF ELEC-CAPITAL	18.0	26.06986	27.41940	28.80791	30.62997	32.55236
COST OF ELEC-FUEL	.0	5.98449	5.98449	5.98449	5.98449	5.98449
COST OF ELEC-OP & MAIN	.0	3.51823	3.51823	3.51823	3.51823	3.51823
TOTAL COST OF ELEC	.0	35.59258	36.92211	38.31063	40.13268	42.05507

Table 9.17 Continued -
BC NO 3 OPEN CYCLE MHD-STEAM BOTTOMING COST OF ELECTRICITY,MILLS/KW.HR
PARAMETRIC POINT NO. 1

ACCOUNT	RATE, PERCENT	FIXED CHARGE RATE, PCT 10.00	14.40	18.00	21.60	25.00
TOTAL DIRECT COSTS,$	.0	786824712.	786824712.	786824712.	736824712.	786824712.
INDIRECT COST,$	51.0	101091141.	101091141.	101091141.	101091141.	101091141.
PROF & OWNER COSTS,$	8.0	62945977.	62945977.	62945977.	62945977.	62945977.
CONTINGENCY COST,$	10.0	78682470.	78682470.	78682470.	78682470.	78682470.
SUB TOTAL,$	.0	1029544288.	1029544288.	1029544288.	1029544288.	1029544288.
ESCALATION COST,$	6.5	308582780.	308582780.	308582780.	308582780.	308582780.
INTREST DURING CONST,$	10.0	380226224.	380226224.	380226224.	380226224.	380226224.
TOTAL CAPITALIZATION,$	.0	1718353280.	1718353280.	1718353280.	1718353280.	1718353280.
COST OF ELEC-CAPITAL	25.0	16.00446	23.04633	28.80791	34.56950	40.01099
COST OF ELEC-FUEL	.0	5.98449	5.98449	5.98449	5.98449	5.98449
COST OF ELEC-OP & MAIN	.0	3.51823	3.51823	3.51823	3.51823	3.51823
TOTAL COST OF ELEC	.0	25.50711	32.54904	38.31063	44.07221	49.51370

ACCOUNT	RATE, PERCENT	FUEL COST, $/10**6 BTU .50	.85	1.50	2.50	1.02
TOTAL DIRECT COSTS,$	.0	786824712.	786824712.	786824712.	786824712.	786824712.
INDIRECT COST,$	51.0	101091141.	101091141.	101091141.	101091141.	101091141.
PROF & OWNER COSTS,$	8.0	62945977.	62945977.	62945977.	62945977.	62945977.
CONTINGENCY COST,$	10.0	78682470.	78682470.	78682470.	78682470.	78682470.
SUB TOTAL,$	.0	1029544288.	1029544288.	1029544288.	1029544288.	1029544288.
ESCALATION COST,$	6.5	308582780.	308582780.	308582780.	308582780.	308582780.
INTREST DURING CONST,$	10.0	380226224.	380226224.	380226224.	380226224.	380226224.
TOTAL CAPITALIZATION,$	.0	1718353280.	1718353280.	1718353280.	1718353280.	1718353280.
COST OF ELEC-CAPITAL	18.0	28.80791	28.80791	28.80791	28.80791	28.80791
COST OF ELEC-FUEL	.0	3.52029	5.98449	10.56086	17.60143	7.18138
COST OF ELEC-OP & MAIN	.0	3.51823	3.51823	3.51823	3.51823	3.51823
TOTAL COST OF ELEC	.0	35.84643	38.31063	42.88700	49.92757	39.50752

ACCOUNT	RATE, PERCENT	CAPACITY FACTOR, PERCENT 12.00	45.00	50.00	65.00	80.00
TOTAL DIRECT COSTS,$	.0	786824712.	786824712.	786824712.	786824712.	786824712.
INDIRECT COST,$	51.0	101091141.	101091141.	101091141.	101091141.	101091141.
PROF & OWNER COSTS,$	8.0	62945977.	62945977.	62945977.	62945977.	62945977.
CONTINGENCY COST,$	10.0	78682470.	78682470.	78682470.	78682470.	78682470.
SUB TOTAL,$	.0	1029544288.	1029544288.	1029544288.	1029544288.	1029544288.
ESCALATION COST,$	6.5	308582780.	308582780.	308582780.	308582780.	308582780.
INTREST DURING CONST,$	10.0	380226224.	380226224.	380226224.	380226224.	380226224.
TOTAL CAPITALIZATION,$	.0	1718353280.	1718353280.	1718353280.	1718353280.	1718353280.
COST OF ELEC-CAPITAL	18.0	156.04287	41.61143	37.45029	28.80791	23.40643
COST OF ELEC-FUEL	.0	5.98449	5.98449	5.98449	5.98449	5.98449
COST OF ELEC-OP & MAIN	.0	3.51823	3.51823	3.51823	3.51823	3.51823
TOTAL COST OF ELEC	.0	165.54559	51.11414	46.95300	38.31063	32.90914

Table 9.18 BC NO 3 OPEN CYCLE MHD-STEAM BOTTOMING

ACCOUNT NO	AUX POWER,MWE	PERC PLANT POW	OPERATION COST	MAINTENANCE COST
4	13.57314	.71983	77.88617	21.25863
7	8.73437	.46321	1693.91193	.00000
8	18.24460	.96756	5.51495	.00000
13	.00000	.00000	13.21918	.00000
14	.00000	.00010	105.28485	.00000
18	13.38680	.70994	.00000	.00000
20	33.34853	1.76857	4261.11108	.00000
21	2.30000	.12198	.00000	.00000
TOTALS	71.37344	3.78513	5634.05585	21.25863

BC NO 3 OPEN CYCLE MHD-STEAM BOTTOMING BASE CASE INPUT

NOMINAL POWER, MWE	1957.0000	NET POWER, MWE	1885.6266
NOM HEAT RATE, BTU/KW-HR	6783.7972	NET HEAT RATE, BTU/KW-HR	7040.5727
OFF DESIGN HEAT RATE	1.0200		
CONDENSER			
DESIGN PRESSURE, IN HG A	2.0000	NUMBER OF SHELLS	5.0000
NUMBER OF TUBES/SHELL	5929.9565	TUBE LENGTH, FT	75.2339
U, BTU/HR-FT2-F	508.8535	TERMINAL TEMP DIFF, F	5.0000
HEAT REJECTION			
DESIGN TEMP, F	51.4000	APPROACH, F	21.6744
RANGE, F	23.0000	OFF DESIGN TEMP, F	77.0000
OFF DESIGN PRES, IN HG A	3.0031	LP TURBINE BLADE LEN, IN	28.5000

1	1957.000	2	.000	3	.514	4	6633.900	5	7.000
6	630.300	7	2.000	8	1397.300	9	5.000	10	2.000
11	1.000	12	369.400	13	1.000	14	4.000	15	1.000
16	2.000	17	234.000	18	3.000	19	5.000	20	3.000
21	.000	22	68500.000	23	.000	24	3550.000	25	400.000
26	9312000.000	27	15000.000	28	24000.000	29	2250000.000	30	.400
31	1.000	32	3635.000	33	.000	34	.700	35	.700
36	5100000.000	37	570.000	38	1.000	39	1.000	40	852000.000
41	157000.000	42	772600.000	43	16500.000	44	1390000.000	45	835000.000
46	1.000	47	.000	48	3.000	49	1.000	50	5.000
51	368.400	52	5.350						

1	269.000	2	144.000	3	16.400	4	96.000	5	18.500
6	163.000	7	350.000	8	159.000	9	3305.000	10	150.000
11	1430.000	12	150.000	13	131.000	14	150.000	15	159.000
16	150.000	17	509.000	18	150.000	19	49.700	20	49.700
21	233.000	22	292.000	23	1.000	24	1.000	25	1.000
26	1.000	27	1.000	28	1.000	29	1.000	30	1.000
31	137.000	32	274.000	33	500.000	34	1225.000	35	3000.000
36	3000.000	37	3000.000	38	3000.000	39	9700000.000	40	.000
41	95000000.000	42	.000	43	230000.000	44	.000	45	780000.000
46	.000	47	5590000.000	48	.000	49	11900000.000	50	.000
51	23800000.000	52	.000	53	530000.000	54	.030	55	.000
56	1.000	57	1.000	58	1.000	59	3650000.000	60	.000
61	12000000.000	62	.000	63	15700000.000	64	.000	65	720000.000
66	1.000	67	.000	68	.000	69	.000	70	.000
71	.000	72	.000	73	.000	74	.000	75	.000
76	.000	77	.000	78	.000	79	725.000	80	2920.000
81	2699.000	82	168.000	83	2397.000	84	111.000	85	1137.000
86	.000	87	.000	88	.000	89	.100	90	.000
91	.000	92	.000	93	.000	94	.100	95	.000
96	.000	97	.000	98	450.000	99	5189.300	100	116.000
101	2.000	102	3540.000	103	1020.000	104	11316.000	105	6720.000
106	2940.000	107	1800.000	108	585.600	109	513.600	110	230.400
111	180.000	112	564.000	113	137.200	114	420.000	115	190.800
116	.000	117	.000	118	.000	119	.000	120	.000
121	.000	122	.000	123	.000	124	.000	125	.000
126	.000	127	.000	128	.000	129	.000	130	.000
131	.000	132	.000	133	.000	134	.000	135	.000
136	.000	137	.000	138	.000	139	.000	140	.000
141	912.000	142	2135.000	143	20.400	144	40.800	145	84.000
146	96.000	147	48000.000	148	12000.000	149	43.000	150	19.100
151	5.000	152	2.500	153	3500.000	154	110.000	155	.000
156	450.000	157	225.000	158	1200.000	159	800.000	160	1200.000
161	1100.000	162	1.000	163	.000	164	.000	165	1.000
166	.000	167	1.000	168	1900000.000	169	3360000.000	170	.000
171	.000	172	4000000.000	173	.000	174	.000	175	2.300
176	.000	177	1.300	178	.413	179	.000	180	.000
181	1.200	182	.000						

Table 9.19 BC NO 3 OPEN CYCLE MHD-STEAM BOTTOMING SUMMARY PLANT RESULTS

PARAMETRIC POINT	1	2	3	4	5	6	7	8
THERMODYNAMIC EFF	.000	.000	.000	.000	.000	.000	.000	.000
POWER PLANT EFF	.435	.471	.459	.515	.535	.000	.000	.000
OVERALL ENERGY EFF	.485	.471	.459	.515	.535	.000	.000	.000
CAP COST MILLION $	1718.353	1347.879	559.961	1437.639	1527.933	.000	.000	.000
CAPITAL COST,$/KWE	911.290	926.874	1007.161	7[illegible]7.632	856.859	.000	.000	.000
COE CAPITAL	28.809	29.301	31.833	24.899	27.097	.000	.000	.000
COE FUEL	5.984	6.157	6.325	5.628	5.421	.000	.000	.000
COE OP & MAIN	3.518	3.748	4.141	3.339	3.218	.000	.000	.000
COST OF ELECTRIC	38.311	39.206	42.306	33.866	35.726	.000	.000	.000
EST TIME OF CONST	7.000	5.402	5.673	6.999	7.004	.000	.000	.000

The thermodynamic states of the gas at various points in the cycle for Base Case 3 are shown in Table 9.7 near the bottom. The points reference in this table refers to the numbered stations on the cycle schematic (Figure 9.2). Table 9.19 presents the efficiency summary for Base Case 3 (Point 1) and its variations. Table 9.20 is the summary detailing the major component material costs and cost of electricity for Base Case 3 points. The same major component groupings are used here as in Base Case 2. Table 9.21 lists the natural resource usage for each Base Case 3 points.

Base Case 3 has a substantially higher capital cost than Base Case 2 due to the cost of the gasifiers and the increased magnet costs (more than twice that of Base Case 2). The increased magnet cost is largely due to the higher MHD pressure ratio used and partly to the increased hydroxyl ion in the gas due to the steam addition in the gasifier. In addition to the increased capital costs, the O&M costs of this point are more than 0.694 mill/MJ (2.5 mills/kWh) higher than that of Base Case 2. Both the high capital cost and O&M can be reduced by better choices of parameters.

The high pressure level of this cycle was chosen to reduce gasifier size and cost. According to Figure 4.32, the gasifier cost penalty is very small compared to the magnet cost penalty. Since 6 or 7 appears to be close to the optimum pressure ratio for a 2700°F (4400°F) combustor temperature (from Base Case 2 results), a substantial improvement in energy cost should result from a reduction in the cycle pressure ratio.

Since it was not necessary for the seed material to collect sulfur in this case, cesium was chosen as the seed material, with a concentration of 1%. The precipitator efficiency was set by the particulate emission standards (99.53%). Due to the high cost of cesium [assumed to be $2.87/kg ($1.30/lb) in the form of pollucite ore], the cost of makeup is rather high. If more efficient precipitators were specified, a lower seed concentration used, or potassium used as the seed, seed makeup costs could be substantially reduced.

Table 9.20 BC NO 3 OPEN CYCLE MHD-STEAM BOTTOMING SUMMARY PLANT RESULTS

	PARAMETRIC POINT		1	2	3	4	5	6	7	8
	TOTAL CAPITAL COST	,M$	1719.35	1347.88	569.96	1487.64	1527.93	.00	.00	.00
P	MHD COMBUSTOR	,M$	.691	.505	.354	.593	.497	.000	.000	.000
L	MHD GENERATOR DUCT	,M$	1.757	1.047	.543	.703	1.396	.000	.000	.000
A	MAGNET & REFRIGERATOR	,M$	164.009	104.540	71.124	84.323	147.356	.000	.000	.000
N	HIGH TEMP HEAT EXCHANGERS	,M$	49.055	29.340	14.900	47.307	46.204	.000	.000	.000
T	SEED RECOVERY SYSTEM	,M$	25.800	18.020	10.610	25.230	24.740	.000	.000	.000
	INVERTER-TRANSFORMER SYSTEM	,M$	65.478	38.294	18.419	71.971	78.939	.000	.000	.000
	COMPRESSOR & DRIVE	,M$	17.490	15.400	7.880	16.200	16.700	.000	.000	.000
	STEAM TURB-GEN	,M$	24.330	17.410	9.810	19.580	17.060	.000	.000	.000
R	TOT MAJOR COMPONENT COST	,M$	349.610	224.556	133.641	265.908	332.891	.000	.000	.000
E	TOT MAJOR COMPONENT COST	,$/KWE	185.408	198.625	236.152	140.785	175.217	.000	.000	.000
S	BALANCE OF PLANT COST	,$/KWE	126.747	134.111	148.817	122.592	118.701	.000	.000	.000
U	SITE LABOR	,$/KWE	105.120	112.601	129.830	95.367	98.464	.000	.000	.000
L	TOTAL DIRECT COST	,$/KWE	417.275	445.337	513.799	358.744	392.382	.000	.000	.000
T	INDIRECT COSTS	,$/KWE	53.611	57.426	65.703	48.637	50.217	.000	.000	.000
	PROF & OWNER COSTS	,$/KWE	33.382	35.627	41.104	28.699	31.391	.000	.000	.000
B	CONTINGENCY COST	,$/KWE	41.727	41.871	44.564	35.870	39.253	.000	.000	.000
R	ESCALATION COST	,$/KWE	163.650	157.010	156.999	141.427	153.928	.000	.000	.000
E	INT DURING CONSTRUCTION	,$/KWE	201.644	199.604	184.992	174.255	189.589	.000	.000	.000
A	TOTAL CAPITALIZATION	,$/KWE	911.290	926.874	1007.161	787.632	856.859	.000	.000	.000
K	COST OF ELEC-CAPITAL	,MILLS/KWE	28.808	29.301	31.839	24.899	27.087	.000	.000	.000
D	COST OF ELEC-FUEL	,MILLS/KWE	5.984	6.157	6.325	5.628	5.421	.000	.000	.000
O	COST OF ELEC-OP&MAIN	,MILLS/KWE	3.518	3.748	4.141	3.339	3.218	.000	.000	.000
W	TOTAL COST OF ELEC	,MILLS/KWE	38.311	39.206	42.306	33.866	35.726	.000	.000	.000
N	COE 0.5 CAP. FACTOR	,MILLS/KWE	46.953	47.996	51.857	41.335	43.852	.000	.000	.000
	COE 0.8 CAP. FACTOR	,MILLS/KWE	32.909	33.712	36.336	29.197	30.647	.000	.000	.000
	COE 1.2XCAP. COST	,MILLS/KWE	44.072	45.066	48.673	38.846	41.144	.000	.000	.000
	COE 1.2XFUEL COST	,MILLS/KWE	39.508	40.437	43.571	34.991	36.810	.000	.000	.000
	COE (CONTINGENCY=0)	,MILLS/KWE	36.139	37.092	40.172	31.973	33.654	.000	.000	.000
	COE (ESCALATION=0)	,MILLS/KWE	31.885	33.155	36.391	28.313	29.682	.000	.000	.000

Table 9.21 BC NO 3 OPEN CYCLE MHD-STEAM BOTTOMING NATURAL RESOURCE REQUIREMENTS

PARAMETRIC POINT	1	2	3	4	5	6	7	8
COAL, LB/KW-HR	.64905	.56731	.68603	.51044	.58792	.00000	.00000	.00000
SORBANT OR SEED, LB/KW-HR	.33756	.34730	.35677	.31750	.30578	.00000	.00000	.00000
TOTAL WATER, GAL/KW-HR	.558	.595	.630	.481	.438	.000	.000	.000
COOLING WATER	.516	.552	.586	.443	.401	.000	.000	.000
GASIFIER PROCESS H2O	.03656	.03752	.03864	.03438	.03312	.00000	.00000	.00000
CONDENSATE MAKE UP	.00493	.00527	.00560	.00423	.00384	.00000	.00000	.00000
WASTE HANDLING SLURRY	.0000	.0000	.0000	.0000	.0000	.0000	.0000	.0000
SCRUBBER WASTE WATER	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
NOX SUPPRESSION	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
TOTAL LAND ACRES/100MWE	91.43	100.11	113.56	85.16	81.13	.00	.00	.00
MAIN PLANT	12.41	16.88	25.58	12.38	12.33	.00	.00	.00
DISPOSAL LAND	55.88	57.50	59.07	52.56	50.62	.00	.00	.00
LAND FOR ACCESS RR	23.14	25.73	29.92	20.22	18.18	.00	.00	.00

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On the negative side, the air and fuel gas must be preheated to high temperatures in order to obtain the desired flame temperature (Table 9.7). Even in the absence of slag and sulfur, the materials problems of these heat exchangers are significant. The fuel gas contains hydrogen which reacts with silicon carbonate at temperatures over 1589°K (2400°F) (Appendix A 9.2). It will probably be necessary to preheat the air to even higher temperatures [∿ 1900°K (2960°F)] if the fuel gas preheat is limited.

Points 2 and 3 are scaled versions of the base case, and the energy costs increase slightly because of this reduction in station size.

In Point 4, the preheat temperature of both the air and fuel gas was increased by 333°K (600°F), with a corresponding 155°K (279°F) increase in combustor temperature. The increased recovery of heat to the MHD plant results in an increase of efficiency of three points to 51.8%; and the increased gas conductivity reduces the magnet cost by $80 million. The total results is that energy costs are reduced to 1.222 mills/MJ (4.4 mills/kWh) below the base case.

Point 5 has an MHD pressure ratio of 15 with the same preheat conditions as Point 4. The efficiency is two points better than that of Point 4, but the equipment cost is $66 /kW more. The result is that energy costs are 0.5 mill/MJ (1.8 mills/kWh) greater than for Point 4.

9.4.3 Base Case 1 and Variations

The detailed accounts listing, cost of electricity summary, and input-output sheets for Base Case 3 are given as Tables 9.22, 9.23, and 9.24, respectively. In this case, 67.03% of the total direct cost are found in Accounts 11, 12, 13, and 15. The cost of some of the major component groupings were $71,200,000 for the superconducting magnet, $17,600,000 for the steam turbine-driven compressor, $26,400,000 for the steam turbine generator, $58,980,000 for the heat recovery steam generator, $212,600,000 for the air heater system, and $62,930,000 for the inverter system. The thermodynamic states of the gas at the various points in the cycle for Base Case 1 are shown in Table 9.9 near the bottom. The point reference in this table refers to the numbered

Table 9.22   8C NO 1 OPEN CYCLE MHD-STEAM BOTTOMING ACCOUNT LISTING

PARAMETRIC POINT NO. 1

ACCOUNT NO. & NAME,	UNIT	AMOUNT	MAT $/UNIT	INS $/UNIT	MAT COST,$	INS COST,$
SITE DEVELOPMENT						
1. 1 LAND COST	ACRE	263.0	1000.00	.00	263000.00	.00
1. 2 CLEARING LAND	ACRE	87.7	.00	600.00	.00	52594.74
1. 3 GRADING LAND	ACRE	263.0	.00	3000.00	.00	789000.00
1. 4 ACCESS RAILROAD	MILE	5.0	115000.00	110000.00	575000.00	550000.00
1. 5 LOOP RAILROAD TRACK	MILE	3.0	120000.00	70000.00	360000.00	210000.00
1. 6 SIDING R R TRACK	MILE	.0	125000.00	80000.00	.00	.00
1. 7 OTHER SITE COSTS	ACRE	.0	.00	.00	534540.10	534540.10
PERCENT TOTAL DIRECT COST IN ACCOUNT 1 = .570 ACCOUNT TOTAL,$					1732540.09	2136134.81
EXCAVATION & PILING						
2. 1 COMMON EXCAVATION	YD3	288000.0	.00	3.00	.00	864000.00
2. 2 PILING	FT	768000.0	6.50	8.50	4992000.00	6528000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 2 = 1.925 ACCOUNT TOTAL,$					4992000.00	7392000.00
PLANT ISLAND CONCRETE						
3. 1 PLANT IS. CONCRETE	YD3	96000.0	70.00	80.00	6720000.00	7680000.00
3. 2 SPECIAL STRUCTURES	YD3	.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 3 = 2.123 ACCOUNT TOTAL,$					6720000.00	7680000.00
HEAT REJECTION SYSTEM						
4. 1 COOLING TOWERS	EACH	29.0	.00	.00	4298000.00	2142000.00
4. 2 CIRCULATING H2O SYS	EACH	1.0	.00	.00	1856966.62	2489957.53
4. 3 SURFACE CONDENSER	FT2	677607.9	.00	.00	3065763.44	474325.53
PERCENT TOTAL DIRECT COST IN ACCOUNT 4 = 2.112 ACCOUNT TOTAL,$					9220730.00	5106283.00
STRUCTURAL FEATURES						
5. 1 STAT. STRUCTURAL ST.	TON	3300.0	650.00	175.00	2145000.00	577500.00
5. 2 SILOS & BUNKERS	TPH	619.7	1800.00	750.00	1115446.92	464769.55
5. 3 CHIMNEY	FT	400.0	.00	.00	435070.92	652606.38
5. 4 STRUCTURAL FEATURES	EACH	1.0	1048000.00	268000.00	1048000.00	268000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 5 = .989 ACCOUNT TOTAL,$					4743517.81	1962875.94
BUILDINGS						
6. 1 STATION BUILDINGS	FT3	8375000.0	.16	.16	1340000.00	1340000.00
6. 2 ADMINSTRATION	FT2	15000.0	16.00	14.00	240000.00	210000.00
6. 3 WAREHOUSE & SHOP	FT2	24000.0	12.00	8.00	288000.00	192000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 6 = .532 ACCOUNT TOTAL,$					1868000.00	1742000.00
FUEL HANDLING & STORAGE						
7. 1 COAL HANDLING SYS.	TPH	649.2	.00	.00	11739704.00	4715334.94
7. 2 DOLOMITE HAND. SYS	TPH	15.6	.00	.00	300824.55	195435.57
7. 3 FUEL OIL HAND. SYS	GAL	2600000.0	.00	.00	290836.01	227826.41
PERCENT TOTAL DIRECT COST IN ACCOUNT 7 = 2.575 ACCOUNT TOTAL,$					12331364.50	5138596.87
FUEL PROCESSING						
8. 1 COAL DRYER & CRUSHER	TPH	.0	.00	.00	.00	.00
8. 2 CARBONIZERS	TPH	619.7	.00	.00	11736526.50	6601796.12
8. 3 GASIFIERS	TPH	.0	.00	.00	.00	796.12
PERCENT TOTAL DIRECT COST IN ACCOUNT 8 = 2.703 ACCOUNT TOTAL,$					11736526.50	6601796.12

Table 9.22 SC NO 1 OPEN CYCLE MHD-STEAM BOTTOMING ACCOUNT LISTING
PARAMETRIC POINT NO. 1
Continued

ACCOUNT NO. & NAME,	UNIT	AMOUNT	MAT $/UNIT	INS $/UNIT	MAT COST,$	INS COST,$
FIRING SYSTEM						
9. 1 CONTAINMENT STEEL	KP	417.1	350.00	159.00	145985.00	66318.90
9. 2 BRICK- DENSE ZRO2	KP	326.3	3305.00	150.00	1078421.48	48945.00
9. 3 BRICK-SILICON CARBIDE	KP	.0	.00	150.00	.00	.00
9. 4 BRICK- MGO	KP	219.0	131.00	150.00	28689.00	32850.00
9. 5 BRICK (INSULATING)	KP	42.6	159.00	150.00	6773.40	6390.00
9. 6 STRUCTURAL STEEL	KP	301.5	509.00	156.00	153463.50	47034.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 9 = .238 ACCOUNT TOTAL,$					1413332.36	201537.90
VAPOR GENERATOR (FIRED)						
10. 1		.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 10 = .000 ACCOUNT TOTAL,$					.00	.00
ENERGY CONVERTER						
11. 1 DUCT INSULATION BRICK	KP	18.1	137.00	274.00	2474.22	4948.44
11. 2 DUCT CONDUCTING BRICK	KP	18.1	500.00	1225.00	9030.00	22123.50
11. 3 DUCT STRUCTURAL STEEL	KP	125.6	3000.00	3000.00	376800.00	376800.00
11. 4 DUCT COOLING TUBES	KP	233.8	3000.00	3000.00	701399.99	701399.99
11. 5 SUPERCONDUCTING MAT	EACH	1.0	29599999.75	7399999.94	29599999.75	7399999.94
11. 6 MAGNET STRUCT STEEL	EACH	1.0	34199999.50	.00	34199999.50	.00
11. 7 NITROGEN LIQUIFIER	EACH	1.0	72900.00	8100.00	72900.00	8100.00
11. 8 HELIUM LIQUIFIER	EACH	1.0	242100.00	26900.00	242100.00	26900.00
11. 9 COMPRESSOR	EACH	1.0	5800000.00	579999.99	5800000.00	579999.99
11.10 COMPRESSOR DRIVE	EACH	1.0	10400000.00	832209.72	10400000.00	832209.72
11.11 STEAM TURB-GEN	EACH	1.0	24900000.00	1509982.81	24900000.00	1509982.81
11.12 FEED WATER HEATERS	PLANT	1.0	603000.00	18090.00	603000.00	18090.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 11 =17.450 ACCOUNT TOTAL,$					106907702.00	11480553.75
COUPLING HEAT EXCHANGER						
12. 1 HIGH TEMP SUPERHEATER	EA	1.0	3392279.97	1696139.98	3392279.97	1696139.98
12. 2 REHEATER & EVAP SECT	EA	1.0	9909360.00	6606239.94	9909360.00	6606239.94
12. 3 ECONOMIZER SECTION	EA	1.0	26166000.00	11214000.00	26166000.00	11214000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 12 = 8.694 ACCOUNT TOTAL,$					39467639.50	19516379.75

Table 9.22 Continued

SC NO 1 OPEN CYCLE MHD-STEAM BOTTOMING ACCOUNT LISTING
PARAMETRIC POINT NO. 1

ACCOUNT NO. & NAME,	UNIT	AMOUNT	MAT $/UNIT	INS $/UNIT	MAT COST,$	INS COST,$
HEAT RECOVERY HEAT EXCH.						
13. 1 CERAMIC TUBING	KP	719.0	2640.00	1008.00	6203520.00	723744.00
13. 2 EXOTIC METAL TUBES	KP	6194.0	11315.00	6720.00	69979144.00	41556480.00
13. 3 STAINLESS STEEL TUBES	KP	6690.0	2940.00	1300.00	19668600.00	12042000.00
13. 4 TUBE CERAMIC COATING	KP	164.0	585.60	513.60	96038.40	84230.40
13. 5 INSULATION FOR REGEN	KP	4393.0	230.40	180.00	1012147.20	790740.00
13. 6 STRUCTURAL STEEL	KP	204.0	564.00	187.20	115056.00	38188.80
13. 7 CONT STEEL REGEN	KP	2034.0	420.00	195.30	975280.00	397627.20
13. 8 CHECKER BRICKS	KP	38080.0	435.60	180.00	16587647.87	6854400.00
13. 9 INSULATION SFA HEATER	KP	20145.0	230.40	180.00	4641407.94	3626100.00
13.10 CONT ST SFA HEATER	KP	4074.0	420.00	190.80	1711080.00	777319.20
13.11 STRU ST SFA HEATER	KP	856.0	564.00	187.20	482784.00	160243.20
13.12 BURNER SFA HEATER	EACH	40.0	33960.00	.00	1358400.00	.00
13.13 HIGH TEMP VALVES	EACH	70.0	79456.80	2760.00	5561975.94	193200.00
13.14 PREHEATER AIR PIPING	TON	5377.0	855.60	568.80	4600561.19	3058437.56
13.15 FUEL GAS PIPING	TON	312.0	816.00	374.80	254592.00	272937.60
13.16 COMB AIR PIPING	TON	2940.0	238.80	250.80	702071.99	737351.99
13.17 RECIRC AIR PIPING	TON	1.0	.00	.00	.00	.00
13.18 PRODUCTS PIPING	TON	3245.0	717.60	369.60	2330764.78	1200460.78
13.19 RECIRC ASPIRATOR	TON	1.0	277200.00	301200.00	277200.00	301200.00
13.20 HEADERS	KP	694.0	912.00	2136.00	632928.00	1482384.00
13.21 SEALS	EACH	7441.0	20.40	40.80	151796.40	303592.90
13.22 CONCRETE	YD3	3775.0	84.00	96.00	317100.00	362400.00
13.23 EXPANSION JOINTS	EACH	1.0	48000.00	12000.00	48000.00	12000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 13 =31.335 ACCOUNT TOTAL,$					[illegible]	74975032.00
WATER TREATMENT						
14. 1 DEMINERALIZER	GPM	342.0	2000.00	560.00	683963.23	191509.71
14. 2 CONDENSATE POLISHING	KWE	1174500.0	1.25	.30	1468124.97	352350.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 14 = .337 ACCOUNT TOTAL,$					2152088.19	543859.70
POWER CONDITIONING						
15. 1 INVERTER SYSTEM	KWE	1167500.0	43.00	4.18	50202500.00	4874312.50
15. 2 FILTERS	KWE	1167500.0	6.00	.73	7005000.00	846437.50
15. 3 STD TRANSFORMERS	KWE	1057466.7	.00	.00	1808194.62	36163.89
PERCENT TOTAL DIRECT COST IN ACCOUNT 15 = 9.547 ACCOUNT TOTAL,$					59015694.50	5756913.87
AUXILIARY MECH EQUIPMENT						
16. 1 BOILER FEED PUMP &DR.	KWE	1115775.0	1.67	.10	1863344.22	111577.50
16. 2 OTHER PUMPS	KWE	955532.5	.83	.12	849660.59	115863.90
15. 3 MISC SERVICE SYS	KWE	1931065.0	1.17	.73	2259346.03	1409677.42
16. 4 AUXILIARY BOILER	PPH	520000.0	4.00	.80	2080000.00	416000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 16 = 1.342 ACCOUNT TOTAL,$					7052358.81	2053118.80
PIPE & FITTINGS						
17. 1 CONVENTIONAL PIPING	TON	4000.0	3000.00	1800.00	12000000.00	7200000.00
17. 2 HIGT TEMP AIR PIPINC	TON	2520.0	450.00	225.00	1134000.00	567000.00
17. 3 LOW TEMP AIR PIPING	TON	117.0	1200.00	800.00	140400.00	93600.00
17. 4 RECIR PRODUCT PIPINC	TON	40.0	1200.00	800.00	48000.00	32000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 17 = 3.127 ACCOUNT TOTAL,$					13322400.00	7892600.00

Table 9.22 PC NO 1 OPEN CYCLE MHD-STEAM BOTTOMING ACCOUNT LISTING
Continued PARAMETRIC POINT NO. 1

ACCOUNT NO. & NAME,	UNIT	AMOUNT	MAT $/UNIT	INS $/UNIT	MAT COST,$	INS COST,$
AUXILIARY ELEC EQUIPMENT						
18. 1 MISC MOTERS,ETC		1544832.0	1.42	.17	2162792.75	262624.93
18. 2 SWITCHGEAR & MCC EAN	KWE	1544852.0	1.95	.45	3012461.31	695183.38
18. 3 CONDUIT,CABLES,TRAYS	FT	6995000.0	1.32	1.36	9220199.87	9499599.87
18. 4 ISOLATED PHASE BUS	FT	570.0	510.00	450.00	290700.00	256500.00
18. 5 LIGHTING & COMMUN	KWE	1931065.0	.35	.43	675872.74	830357.94
PERCENT TOTAL DIRECT COST IN ACCOUNT 18 = 3.966 ACCOUNT TOTAL,$					15362026.50	11544265.87
CONTROL, INSTRUMENTATION						
19. 1 COMPUTER	EACH	1.0	726000.00	16500.00	726000.00	16500.00
19. 2 OTHER CONTROLS	EACH	1.0	1391500.00	835000.00	1391500.00	835000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 19 = .438 ACCOUNT TOTAL,$					2117500.00	851500.00
PROCESS WASTE SYSTEMS						
20. 1 BOTTOM ASH	TPH	56.1	5284999.37	1321249.84	5284999.37	1321249.84
20. 2 DRY ASH	TPH	6.2	454957.83	113739.46	454957.83	113739.46
20. 3 WET SLURRY	TPH	.0	.00	.00	.00	.00
20. 4 ONSITE DISPOSAL	ACRS	271.8	6491.62	9637.32	1764492.31	2619528.37
20. 5 SEED TREATMENT	EACH	1.0	16441800.00	8098199.94	16441800.00	8098199.94
PERCENT TOTAL DIRECT COST IN ACCOUNT 20 = 5.321 ACCOUNT TOTAL,$					23946249.50	12152717.50
STACK GAS CLEANING						
21. 1 PRECIPITATOR	EACH	1.0	24080000.00	3610000.00	24080000.00	3610000.00
21. 2 SCRUBBER	KWE	.0	15.68	6.76	.00	.00
21. 3 MISC STEEL & DUCTS		1.0	4300000.00	.00	4300000.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 21 = 4.715 ACCOUNT TOTAL,$					28380000.00	3610000.00
TOTAL DIRECT COSTS,$					490038748.00	138338156.00

Table 9.23

SC NO 1 OPEN CYCLE MHD-STEAM BOTTOMING COST OF ELECTRICITY,MILLS/KW.HR

PARAMETRIC POINT NO. 1

ACCOUNT	RATE, PERCENT	LABOR RATE, $/HR 6.00	8.50	10.60	15.00	21.50
TOTAL DIRECT COSTS,$	.0	596695248.	641114624.	678426896.	756605000.	872095376.
INDIRECT COST,$	51.0	54369715.	77023197.	96052458.	135923290.	194923380.
PROF & OWNER COSTS,$	8.0	47735619.	51289169.	54274151.	60529400.	69767630.
CONTINGENCY COST,$	11.0	65636476.	70522608.	74626958.	83226549.	95930490.
SUB TOTAL,$	.0	764436648.	839949552.	903380448.	1036283232.	1232616848.
ESCALATION COST,$	6.5	267657094.	294096924.	316306376.	362840476.	431584024.
INTREST DURING CONST,$	10.0	341105352.	374800564.	403104568.	462408176.	550015760.
TOTAL CAPITALIZATION,$	.0	1373199088.	1508947032.	1622791376.	1861531872.	2214216608.
COST OF ELEC-CAPITAL	18.0	22.02956	24.20570	26.03365	29.86365	35.52160
COST OF ELEC-FUEL	.0	6.04185	6.04185	6.04185	6.04185	6.04185
COST OF ELEC-OP & MAIN	.0	.68771	.68771	.68771	.68771	.69771
TOTAL COST OF ELEC	.0	28.75912	30.93526	32.76321	36.59321	42.25116

ACCOUNT	RATE, PERCENT	CONTINGENCY, PERCENT -5.00	.00	11.00	5.00	20.00
TOTAL DIRECT COSTS,$	.0	678426896.	678426896.	678426896.	678426896.	678426896.
INDIRECT COST,$	51.0	96052458.	96052458.	96052458.	96052458.	96052458.
PROF & OWNER COSTS,$	8.0	54274151.	54274151.	54274151.	54274151.	54274151.
CONTINGENCY COST,$	20.0	-33921344.	0.	74626958.	33921344.	135685378.
SUB TOTAL,$	.0	794832152.	828753496.	903380448.	862674840.	964438872.
ESCALATION COST,$	6.5	278299664.	290176764.	316306376.	302053860.	337685152.
INTREST DURING CONST,$	10.0	354669372.	369904684.	403104568.	384940996.	430349936.
TOTAL CAPITALIZATION,$	.0	1427800176.	1488734926.	1622791376.	1549669680.	1732473952.
COST OF ELEC-CAPITAL	18.0	22.90550	23.88305	26.03365	24.86059	27.79323
COST OF ELEC-FUEL	.0	6.04185	6.04185	6.04185	6.04185	6.04185
COST OF ELEC-OP & MAIN	.0	.68771	.68771	.68771	.68771	.68771
TOTAL COST OF ELEC	.0	29.63506	30.61261	32.76321	31.59015	34.52279

ACCOUNT	RATE, PERCENT	ESCALATION RATE, PERCENT 5.00	6.50	8.00	10.00	.00
TOTAL DIRECT COSTS,$	.0	678426896.	678426896.	[illegible]	678426896.	678426896.
INDIRECT COST,$	51.0	96052458.	96052458.	[illegible]2458.	96052458.	96052458.
PROF & OWNER COSTS,$	8.0	54274151.	54274151.	54274151.	54274151.	54274151.
CONTINGENCY COST,$	11.0	74626958.	74626958.	74626958.	74626958.	74626958.
SUB TOTAL,$	.0	903380448.	903380448.	903380448.	903380448.	903380448.
ESCALATION COST,$	.0	235602156.	316306376.	402089364.	524839296.	0.
INTREST DURING CONST,$	10.0	380784640.	403104568.	426572468.	459734240.	314073988.
TOTAL CAPITALIZATION,$	.0	1519767232.	1622791376.	1732042272.	1887953984.	1217454432.
COST OF ELEC-CAPITAL	18.0	24.38086	26.03365	27.78631	30.28752	19.53103
COST OF ELEC-FUEL	.0	6.04185	6.04185	6.04185	6.04185	6.04185
COST OF ELEC-OP & MAIN	.0	.68771	.68771	.68771	.68771	.68771
TOTAL COST OF ELEC	.0	31.11044	32.76321	34.51587	37.01708	26.26059

ACCOUNT	RATE, PERCENT	INT DURING CONST,PERCENT 6.00	9.00	10.00	12.50	15.00
TOTAL DIRECT COSTS,$	.0	678426896.	678426896.	678426896.	678426896.	678426896.
INDIRECT COST,$	51.0	96052458.	96052458.	96052458.	96052458.	96052458.
PROF & OWNER COSTS,$	8.0	54274151.	54274151.	54274151.	54274151.	54274151.
CONTINGENCY COST,$	11.0	74626958.	74626958.	74626958.	74626958.	74626958.
SUB TOTAL,$	.0	903380448.	903380448.	903380448.	903380448.	903380448.
ESCALATION COST,$	6.5	316306376.	316306376.	316306376.	316306376.	316306376.
INTREST DURING CONST,$	15.0	229041390.	313796748.	403104568.	521482644.	647781400.
TOTAL CAPITALIZATION,$	.0	1448729192.	1533487[illegible]	1622791376.	1741169456.	1867468200.
COST OF ELEC-CAPITAL	18.0	23.24124	[illegible]	26.03365	27.93273	29.95888
COST OF ELEC-FUEL	.0	6.04185	[illegible]	6.04185	6.04185	6.04185
COST OF ELEC-OP & MAIN	.0	.68771	[illegible]	.68771	.68771	.69771
TOTAL COST OF ELEC	.0	29.97080	31.[illegible]49	32.76321	34.66229	36.68844

Table 9.23 Continued -

BC NO 1 OPEN CYCLE MHD-STEAM BOTTOMING COST OF ELECTRICITY,MILLS/KW.HR

PARAMETRIC POINT NO. 1

ACCOUNT	RATE, PERCENT	FIXED CHARGE RATE, PCT 10.00	14.40	18.00	21.60	25.00
TOTAL DIRECT COSTS,$	.0	678426896.	678426896.	678426896.	676426896.	678426896.
INDIRECT COST,$	51.0	96052459.	96052458.	96052459.	96052458.	96052458.
PROF & OWNER COSTS,$	8.0	54274151.	54274151.	54274151.	54274151.	54274151.
CONTINGENCY COST,$	11.0	74626958.	74626958.	74626958.	74626958.	74626958.
SUB TOTAL,$	.0	903380448.	903380448.	903380448.	903380448.	903380448.
ESCALATION COST,$	6.5	316306376.	316306376.	316306376.	316306376.	316306376.
INTREST DURING CONST,$	10.0	403104568.	403104568.	403104568.	403104568.	403104568.
TOTAL CAPITALIZATION,$	.0	1622791376.	1622791376.	1622791376.	1622791376.	1622791376.
COST OF ELEC-CAPITAL	25.0	14.46314	20.82592	26.03365	31.24038	36.15785
COST OF ELEC-FUEL	.0	6.04185	6.04185	6.04185	6.04185	6.04185
COST OF ELEC-OP & MAIN	.0	.68771	.68771	.68771	.68771	.68771
TOTAL COST OF ELEC	.0	21.19270	27.55648	32.76321	37.96994	42.88741

ACCOUNT	RATE, PERCENT	FUEL COST, $/10**6 BTU .50	.85	1.50	2.50	1.02
TOTAL DIRECT COSTS,$	.0	678425896.	678426896.	678426896.	678426896.	678426896.
INDIRECT COST,$	51.0	96052458.	96052458.	96052458.	96052458.	96052458.
PROF & OWNER COSTS,$	8.0	54274151.	54274151.	54274151.	54274151.	54274151.
CONTINGENCY COST,$	11.0	74626958.	74626958.	74626958.	74626958.	74626958.
SUB TOTAL,$	.0	903380448.	903380448.	903380448.	903380448.	903380448.
ESCALATION COST,$	6.5	316306376.	316306376.	316306376.	316306376.	316306376.
INTREST DURING CONST,$	10.0	403104568.	403104568.	403104568.	403104568.	403104568.
TOTAL CAPITALIZATION,$	.0	1622791376.	1622791376.	1622791376.	1622791376.	1622791376.
COST OF ELEC-CAPITAL	18.0	26.03365	26.03365	26.03365	26.03365	26.03365
COST OF ELEC-FUEL	.0	3.55403	6.04185	10.66209	17.77016	7.25022
COST OF ELEC-OP & MAIN	.0	.68771	.68771	.68771	.68771	.68771
TOTAL COST OF ELEC	.0	30.27539	32.76321	37.38345	44.49151	33.97158

ACCOUNT	RATE, PERCENT	CAPACITY FACTOR, PERCENT 12.00	45.00	50.00	65.00	80.00
TOTAL DIRECT COSTS,$	.0	678426896.	678426896.	678426896.	678426896.	678426896.
INDIRECT COST,$	51.0	96052458.	96052458.	96052458.	96052458.	96052458.
PROF & OWNER COSTS,$	8.0	54274151.	54274151.	54274151.	54274151.	54274151.
CONTINGENCY COST,$	11.0	74626958.	74626958.	74626958.	74626958.	74626958.
SUB TOTAL,$	.0	903380448.	903380448.	903380448.	903380448.	903380448.
ESCALATION COST,$	6.5	316306376.	316306376.	316306376.	316306376.	316306376.
INTREST DURING CONST,$	10.0	403104568.	403104568.	403104568.	403104568.	403104568.
TOTAL CAPITALIZATION,$	.0	1622791376.	1622791376.	1622791376.	1622791376.	1622791376.
COST OF ELEC-CAPITAL	18.0	141.01559	37.60416	33.84374	26.03365	21.15234
COST OF ELEC-FUEL	.0	6.04185	6.04185	6.04185	6.04185	6.04185
COST OF ELEC-OP & MAIN	.0	.68772	.68771	.68771	.68771	.68771
TOTAL COST OF ELEC	.0	147.74516	44.33372	40.57330	32.76321	27.88190

Table 9.24 BC NO 1 OPEN CYCLE MHD-STEAM BOTTOMING

ACCOUNT NO	AUX POWER,MWE	PERC PLANT FOR	OPERATION COST	MAINTENANCE COST
4	15.77032	.60021	90.37225	24.53256
7	6.19669	.31447	.00000	.00000
8	.80560	.04988	.00000	.00000
13	.00000	.00000	147.02082	.00000
14	.00000	.00000	31.93320	.00000
18	15.35470	.77322	.00000	.00000
20	21.54431	1.05323	607.42067	.00000
21	2.50000	.12307	.00000	.00000
TOTALS	62.17162	3.15507	1355.14305	24.53256

BC NO 1 OPEN CYCLE MHD-STEAM BOTTOMING BASE CASE INPUT

NOMINAL POWER, MWE	2032.7000	NET POWER, MWE	1970.5284
NOM HEAT RATE, BTU/KW-HR	6890.6577	NET HEAT RATE, BTU/KW-HR	7108.0630
OFF DESIGN HEAT RATE	1.0370		
CONDENSER			
DESIGN PRESSURE, IN HG A	2.0000	NUMBER OF SHELLS	5.0000
NUMBER OF TUBES/SHELL	6830.5931	TUBE LENGTH, FT	75.2339
U, BTU/HR-FT2-F	608.8535	TERMINAL TEMP DIFF, F	5.0000
HEAT REJECTION			
DESIGN TEMP, F	51.4000	APPROACH, F	21.6744
RANGE, F	23.0000	OFF DESIGN TEMP, F	77.0000
OFF DESIGN PRES, IN HG A	2.9632	LP TURBINE BLADE LEN, IN	31.5000

1	2032.700	2	.000	3	.519	4	6577.700	5	8.000
6	865.200	7	2.000	8	1622.000	9	5.000	10	3.000
11	1.000	12	303.300	13	1.000	14	.000	15	.000
16	2.000	17	263.000	18	3.000	19	5.000	20	3.000
21	.000	22	96000.000	23	.000	24	3300.000	25	400.000
26	8375000.000	27	15000.000	28	24000.000	29	2600000.000	30	.500
31	1.000	32	4000.000	33	1.000	34	.800	35	.800
36	6985000.000	37	570.000	38	1.000	39	1.000	40	1048000.000
41	260000.000	42	726000.000	43	16500.000	44	1391500.000	45	835000.000
46	5.000	47	1.000	48	3.000	49	1.000	50	3.000
51	309.300	52	5.350						

1	417.100	2	326.300	3	.000	4	219.000	5	42.600
6	301.500	7	350.000	8	159.000	9	3305.000	10	150.000
11	.000	12	150.000	13	131.000	14	150.000	15	159.000
16	150.000	17	509.000	18	156.000	19	18.060	20	18.060
21	125.600	22	233.800	23	1.000	24	1.000	25	1.000
26	1.000	27	1.000	28	1.000	29	1.000	30	1.000
31	137.000	32	274.000	33	500.000	34	1225.000	35	3000.000
36	3000.000	37	3900.000	38	3000.000	39	37000000.000	40	.000
41	38000000.000	42	.000	43	81000.000	44	.000	45	269000.000
46	.000	47	5800000.000	48	.000	49	10400000.000	50	.000
51	24900000.000	52	.000	53	603000.000	54	.030	55	.000
56	1.000	57	1.000	58	1.000	59	3141000.000	60	.000
61	13763000.000	62	.000	63	1750000.000	64	.000	65	520000.000
66	1.000	67	.000	68	.000	69	.000	70	.000
71	.000	72	.000	73	.000	74	.000	75	.000
76	.000	77	.000	78	.000	79	718.000	80	6194.000
81	6690.000	82	164.000	83	4393.000	84	204.000	85	2084.000
86	38080.000	87	20145.000	88	4074.000	89	856.000	90	40.000
91	70.000	92	5377.000	93	312.000	94	2940.000	95	1.000
96	3248.000	97	1.000	98	694.000	99	7441.000	100	3775.000
101	1.000	102	8640.000	103	1000.000	104	11316.000	105	6720.000
106	2940.000	107	1300.000	108	525.600	109	513.600	110	230.400
111	180.000	112	564.000	113	107.200	114	420.000	115	190.800
116	435.600	117	180.000	118	230.400	119	180.000	120	420.000
121	190.800	122	564.000	123	187.200	124	33960.000	125	.000
126	79456.800	127	2760.000	128	855.600	129	568.800	130	816.000
131	874.800	132	238.800	133	250.800	134	.000	135	.000
136	717.600	137	369.600	138	277200.000	139	.000	140	301200.000
141	912.000	142	2136.000	143	20.400	144	40.800	145	301200.000
146	96.000	147	49003.000	148	12000.000	149	43.000	150	16.700
151	6.000	152	2.000	153	2520.000	154	117.000	155	40.000
156	450.000	157	225.000	158	1200.000	159	800.000	160	1200.000
161	800.000	162	1.000	163	24540000.000	164	.000	165	1.000
166	.000	167	1.000	168	24080000.000	169	3610000.000	170	.000
171	.000	172	4300000.000	173	.000	174	14.190	175	2.500
176	7.430	177	.133	178	.575	179	9.018	180	9.702
181	1.200	182	.000						

Table 9.25 BC NO 1 OPEN CYCLE MHD-STEAM BOTTOMING SUMMARY PLANT RESULTS

PARAMETRIC POINT	1	2	3	4	5	6	7	8
THERMODYNAMIC EFF	.000	.000	.000	.000	.000	.000	.000	.000
POWER PLANT EFF	.480	.471	.472	.480	.498	.492	.480	.491
OVERALL ENERGY EFF	.480	.471	.472	.480	.488	.492	.480	.491
CAP COST MILLION $	1622.791	978.623	527.627	1622.791	1472.847	1464.393	1470.137	1440.568
CAPITAL COST,$/KWE	823.531	835.333	689.225	823.531	763.317	755.265	760.831	748.941
COE CAPITAL	26.034	26.497	28.110	26.034	24.130	23.876	24.052	23.676
COE FUEL	6.042	6.161	6.148	6.042	5.946	5.892	6.041	5.907
COE OP & MAIN	.688	.893	1.189	.688	.834	.819	.741	.780
COST OF ELECTRIC	32.763	33.461	35.447	32.763	30.911	30.587	30.834	30.363
EST TIME OF CONST	8.000	7.305	6.486	8.000	7.965	7.972	7.970	7.963

PARAMETRIC POINT	9	10	11	12	13	14	15	16
THERMODYNAMIC EFF	.000	.000	.000	.000	.000	.000	.000	.000
POWER PLANT EFF	.480	.491	.460	.480	.443	.535	.496	.503
OVERALL ENERGY EFF	.480	.481	.460	.480	.443	.535	.496	.503
CAP COST MILLION $	1623.214	1621.220	1651.779	1268.127	1488.228	1638.479	1634.665	1728.056
CAPITAL COST,$/KWE	823.785	822.608	842.511	648.700	776.144	827.198	826.090	876.811
COE CAPITAL	26.042	26.004	26.634	20.507	24.536	26.150	26.115	27.718
COE FUEL	6.044	6.035	6.308	6.047	6.547	5.426	5.854	5.771
COE OP & MAIN	.606	.921	.554	.649	.576	.739	.627	.693
COST OF ELECTRIC	32.692	32.961	33.496	27.203	31.658	32.315	32.596	34.182
EST TIME OF CONST	8.000	8.000	8.000	7.987	7.965	8.002	8.004	7.997

PARAMETRIC POINT	17	18	19	20	21	22	23	24
THERMODYNAMIC EFF	.000	.000	.000	.000	.000	.000	.000	.000
POWER PLANT EFF	.473	.000	.000	.000	.000	.000	.000	.000
OVERALL ENERGY EFF	.473	.000	.000	.000	.000	.000	.000	.000
CAP COST MILLION $	1599.229	.000	.000	.000	.000	.000	.000	.000
CAPITAL COST,$/KWE	823.098	.000	.000	.000	.000	.000	.000	.000
COE CAPITAL	26.020	.000	.000	.000	.000	.000	.000	.000
COE FUEL	6.129	.000	.000	.000	.000	.000	.000	.000
COE OP & MAIN	.695	.000	.000	.000	.000	.000	.000	.000
COST OF ELECTRIC	32.844	.000	.000	.000	.000	.000	.000	.000
EST TIME OF CONST	7.981	.000	.000	.000	.000	.000	.000	.000

stations on the cycle schematic in Figure 9.3. Table 9.25 presents the efficiency summary for Base Case 1 (Point 1) and its variations. Table 9.26 is made up of the summary detailing the major component material cost and the cost of electricity for all Base Case 1 parameteric points. Table 9.27 gives the natural resource summary for all Base Case 1 points.

Point 1 (Base Case 1) is a 1971 MW plant burning the dried bituminous coal. The efficiency of Base Case 1 is 1.1 points higher than that of Base Case 2, but its capital cost is \$106/kW (∿ 24%) more. This is due largely to the additional costs of the separate air preheaters and carbonizers and the fact that recirculation requires the handling of larger preheat flows. These add about \$40/kW to the direct cost of the plant, plus the escalation, contingency and interest during construction of these components. Some of the difference in cost is the construction time for this plant. The estimated time is 8 years; while for Base Case 2 it is 7. For the standard assumptions, the energy cost for Base Case 1 is 9.11 mills/MJ (32.8 mills/kWh) or nearly 1.39 mills/MJ (5 mills/kWh) higher than for Base Case 2.

Points 2 and 3 are again scaled down versions of the base case and (as in Base Case 2) the capital costs increase modestly at first (as the output goes down from 1971 to 593 MW).

Point 4 was eliminated when it proved to be identical to the base case because of the need to preoxidize the bituminous coal.

Points 5 and 6 use the subbituminous coal with 20% and 16% moisture, respectively. The efficiencies are slightly higher than those of the base case because of the reduced energy and power requirements of the seed treatment system. The very low quality of the fuel gas produced by carbonizing this coal made it impossible to preheat the air to the levels originally specified. We decided, therefore, to use the gas to reduce the size and technical uncertainty of the heat recovery exchangers. The preheat stream was heated to a lower temperature [1340°K (1970°F)] by heat transferred from the MHD exhaust. With the combustion air for the gapor heated to the same temperature, the preheat stream could be heated to about 1700°K (2600°F) in the separate heater. The resultant reduction

Table 9.26 BC NO 1 OPEN CYCLE MHD-STEAM BOTTOMING SUMMARY PLANT RESULTS

	PARAMETRIC POINT		1	2	3	4	5	6	7	8
	TOTAL CAPITAL COST	,M$	1622.79	[illegible]78.62	527.63	1622.79	1472.85	1464.39	1470.14	1440.57
P	MHD COMBUSTOR	,M$	1.413	1.018	.716	1.413	1.411	1.396	1.413	1.413
L	MHD GENERATOR DUCT	,M$	1.090	.750	.403	1.090	.676	.855	.879	1.061
A	MAGNET & REFRIGERATOR	,M$	64.115	[illegible]7.515	22.140	64.115	50.352	49.546	63.256	49.615
N	HIGH TEMP HEAT EXCHANGERS	,M$	137.607	[illegible]3.689	43.395	137.607	116.686	115.980	95.663	102.442
T	SEED RECOVERY SYSTEM	,M$	44.822	30.235	19.321	44.822	29.855	31.689	32.669	32.103
	INVERTER-TRANSFORMER SYSTEM	,M$	57.208	33.580	16.440	57.208	48.015	48.873	47.633	46.864
	COMPRESSOR & DRIVE	,M$	16.200	11.550	7.400	16.200	15.700	15.700	15.200	15.400
	STEAM TURB-GEN	,M$	25.503	19.423	12.218	25.503	27.620	27.630	33.720	33.490
R	TOT MAJOR COMPONENT COST	,M$	347.957	[illegible]17.760	122.113	347.957	290.515	291.653	294.434	282.388
E	TOT MAJOR COMPONENT COST	,$/KWE	176.581	1[illegible]5.876	205.601	176.581	150.562	150.424	152.377	146.612
S	BALANCE OF PLANT COST	,$/KWE	72.129	77.920	93.127	72.129	73.757	73.622	79.325	77.723
U	SITE LABOR	,$/KWE	95.577	104.677	119.544	95.577	92.241	91.057	87.559	88.802
L	TOTAL DIRECT COST	,$/KWE	344.287	368.542	418.872	344.287	318.561	315.103	319.261	313.336
T	INDIRECT COSTS	,$/KWE	48.745	[illegible]3.385	61.171	48.745	47.043	46.439	44.655	45.289
	PROF & OWNER COSTS	,$/KWE	27.543	29.483	33.510	27.543	25.485	25.209	25.541	25.057
B	CONTINGENCY COST	,$/KWE	37.872	[illegible]7.579	39.733	37.872	34.930	34.572	35.023	34.352
R	ESCALATION COST	,$/KWE	160.519	154.101	151.940	160.519	148.399	146.906	147.969	145.587
E	INT DURING CONSTRUCTION	,$/KWE	204.567	1[illegible]1.642	183.999	204.567	188.899	187.040	188.382	185.309
A	TOTAL CAPITALIZATION	,$/KWE	823.531	[illegible]35.333	889.225	823.531	763.317	755.269	760.831	748.941
K	COST OF ELEC-CAPITAL	,MILLS/KWE	26.034	[illegible]6.407	28.110	26.034	24.130	23.876	24.052	23.676
D	COST OF ELEC-FUEL	,MILLS/KWE	6.042	6.161	6.148	6.042	5.946	5.892	6.041	5.907
O	COST OF ELEC-OP&MAIN	,MILLS/KWE	.688	.893	1.189	.688	.834	.819	.741	.780
W	TOTAL COST OF ELEC	,MILLS/KWE	32.763	33.461	35.447	32.763	30.911	30.587	30.834	30.363
N	COE 0.5 CAP. FACTOR	,MILLS/KWE	40.573	41.383	43.880	40.573	38.150	37.750	38.049	37.465
	COE 0.8 CAP. FACTOR	,MILLS/KWE	27.882	[illegible]8.510	30.17[illegible]	27.882	26.386	26.110	26.324	25.923
	COE 1.2XCAP. COST	,MILLS/KWE	37.970	[illegible]8.743	41.069	37.970	35.737	35.362	35.644	35.098
	COE 1.2XFUEL COST	,MILLS/KWE	33.972	34.694	36.677	33.972	32.100	31.765	32.042	31.544
	COE (CONTINGENCY=0)	,MILLS/KWE	30.613	31.412	33.429	30.613	28.932	28.628	28.849	28.417
	COE (ESCALATION=0)	,MILLS/KWE	26.261	27.353	29.576	26.261	24.906	24.641	24.845	24.472

	PARAMETRIC POINT		9	10	11	12	13	14	15	16
	TOTAL CAPITAL COST	,M$	1623.21	1621.22	1651.78	1268.13	1488.23	1638.48	1634.67	1728.06
P	MHD COMBUSTOR	,M$	1.451	.929	.929	.900	1.413	1.173	1.138	.984
L	MHD GENERATOR DUCT	,M$	1.090	1.090	1.090	.703	1.013	1.149	1.240	2.667
A	MAGNET & REFRIGERATOR	,M$	64.115	64.115	64.115	41.992	62.315	70.140	98.844	148.993
N	HIGH TEMP HEAT EXCHANGERS	,M$	137.607	137.607	137.607	69.477	105.153	148.836	115.449	104.070
T	SEED RECOVERY SYSTEM	,M$	44.862	44.881	54.557	36.989	43.152	45.226	44.747	44.412
	INVERTER-TRANSFORMER SYSTEM	,M$	57.208	57.208	57.208	60.579	52.087	73.353	63.984	67.973
	COMPRESSOR & DRIVE	,M$	16.200	16.200	16.200	15.800	15.800	18.200	17.500	18.300
	STEAM TURB-GEN	,M$	25.503	25.503	25.503	25.103	27.110	18.400	25.103	21.803
R	TOT MAJOR COMPONENT COST	,M$	348.035	347.533	357.209	251.543	308.043	376.[illegible]77	368.006	409.201
E	TOT MAJOR COMPONENT COST	,$/KWE	176.629	176.338	182.199	128.675	160.651	190.[illegible]	185.974	207.628
S	BALANCE OF PLANT COST	,$/KWE	72.159	72.052	71.741	68.619	74.295	68.[illegible]	70.447	70.014
U	SITE LABOR	,$/KWE	95.606	95.498	98.131	74.537	90.422	90.[illegible]9	90.929	92.831
L	TOTAL DIRECT COST	,$/KWE	344.393	343.889	352.071	271.831	325.368	348.300	347.351	370.473
T	INDIRECT COSTS	,$/KWE	48.759	48.704	50.047	38.014	46.115	45.925	46.374	47.344
	PROF & OWNER COSTS	,$/KWE	27.551	27.511	28.166	21.746	26.029	27.864	27.788	29.638
B	CONTINGENCY COST	,$/KWE	37.883	37.828	38.728	29.867	35.676	38.321	38.222	40.743
R	ESCALATION COST	,$/KWE	160.568	160.339	164.218	126.323	150.889	161.260	161.064	170.872
E	INT DURING CONSTRUCTION	,$/KWE	204.630	204.337	209.281	160.919	192.066	205.528	205.290	217.742
A	TOTAL CAPITALIZATION	,$/KWE	823.785	822.608	842.511	648.700	776.144	827.198	826.090	876.811
K	COST OF ELEC-CAPITAL	,MILLS/KWE	26.042	26.004	26.634	20.507	24.536	26.150	826.090	876.811
D	COST OF ELEC-FUEL	,MILLS/KWE	6.044	6.035	6.308	6.047	6.547	5.426	5.854	5.771
O	COST OF ELEC-OP&MAIN	,MILLS/KWE	.606	.921	.554	.649	.576	.739	.627	.693
W	TOTAL COST OF ELEC	,MILLS/KWE	32.692	32.961	33.436	27.203	31.658	32.315	32.596	34.182
N	COE 0.5 CAP. FACTOR	,MILLS/KWE	40.505	40.762	41.486	33.355	39.019	40.160	40.431	42.498
	COE 0.8 CAP. FACTOR	,MILLS/KWE	27.809	28.085	28.502	23.358	27.058	27.412	27.700	28.985
	COE 1.2XCAP. COST	,MILLS/KWE	37.900	38.162	38.623	31.304	36.565	37.545	37.819	39.726
	COE 1.2XFUEL COST	,MILLS/KWE	33.901	34.168	34.758	28.412	32.967	33.400	33.767	35.337
	COE (CONTINGENCY=0)	,MILLS/KWE	30.541	[illegible]0.813	31.297	25.509	29.637	30.139	30.425	31.869
	COE (ESCALATION=0)	,MILLS/KWE	26.188	[illegible]6.466	26.843	22.089	25.552	25.782	26.071	27.261

Table 9.26 3C NO 1 OPEN CYCLE MHD-STEAM BOTTOMING SUMMARY PLANT RESULTS
Continued

	PARAMETRIC POINT		17	18	19	20	21	22	23	24
	TOTAL CAPITAL COST	,M$	1599.23	.00	.00	.00	.00	.00	.00	.00
P	MHD COMBUSTOR	,M$	1.413	.000	.000	.000	.000	.000	.000	.000
L	MHD GENERATOR DUCT	,M$	1.090	.000	.000	.000	.000	.000	.000	.000
A	MAGNET & REFRIGERATOR	,M$	64.115	.000	.000	.000	.000	.000	.000	.000
N	HIGH TEMP HEAT EXCHANGERS	,M$	137.607	.000	.000	.000	.000	.000	.000	.000
T	SEED RECOVERY SYSTEM	,M$	44.922	.000	.000	.000	.000	.000	.000	.000
	INVERTER-TRANSFORMER SYSTEM	,M$	57.232	.000	.000	.000	.000	.000	.000	.000
	COMP ESSOR & DRIVE	,M$	16.200	.000	.000	.000	.000	.000	.000	.000
	STEAM TURB-GEN	,M$	25.703	.000	.000	.000	.000	.000	.000	.000
R	TOT MAJOR COMPONENT COST	,M$	348.192	.000	.000	.000	.000	.000	.000	.000
E	TOT MAJOR COMPONENT COST	,$/KWE	179.204	.000	.000	.000	.000	.000	.000	.000
S	BALANCE OF PLANT COST	,$/KWE	69.919	.000	.000	.000	.000	.000	.000	.000
U	SITE LABOR	,$/KWE	95.565	.000	.000	.000	.000	.000	.000	.000
L	TOTAL DIRECT COST	,$/KWE	344.689	.000	.000	.000	.000	.000	.000	.000
T	INDIRECT COSTS	,$/KWE	48.738	.000	.000	.000	.000	.000	.000	.000
	PROF & OWNER COSTS	,$/KWE	27.575	.000	.000	.000	.000	.000	.000	.000
B	CONTINGENCY COST	,$/KWE	37.850	.000	.000	.000	.000	.000	.000	.000
R	ESCALATION COST	,$/KWE	160.203	.000	.000	.000	.000	.000	.000	.000
E	INT DURING CONSTRUCTION	,$/KWE	204.039	.000	.000	.000	.000	.000	.000	.000
A	TOTAL CAPITALIZATION	,$/KWE	823.098	.000	.000	.000	.000	.000	.000	.000
K	COST OF ELEC-CAPITAL	,MILLS/KWE	26.020	.000	.000	.000	.000	.000	.000	.000
D	COST OF ELEC-FUEL	,MILLS/KWE	6.129	.000	.000	.000	.000	.000	.000	.000
O	COST OF ELEC-OP&MAIN	,MILLS/KWE	.695	.000	.000	.000	.000	.000	.000	.000
W	TOTAL COST OF ELEC	,MILLS/KWE	32.849	.000	.000	.000	.000	.000	.000	.000
N	COE 0.5 CAP. FACTOR	,MILLS/KWE	40.650	.000	.000	.000	.000	.000	.000	.000
	COE 0.8 CAP. FACTOR	,MILLS/KWE	27.965	.000	.000	.000	.000	.000	.000	.000
	COE 1.2XCAP. COST	,MILLS/KWE	38.040	.000	.000	.000	.000	.000	.000	.000
	COE 1.2XFUEL COST	,MILLS/KWE	34.069	.000	.000	.000	.000	.000	.000	.000
	COE (CONTINGENCY=0)	,MILLS/KWE	30.697	.000	.000	.000	.000	.000	.000	.000
	COE (ESCALATION=0)	,MILLS/KWE	26.357	.000	.000	.000	.000	.000	.000	.000

Table 9.27 BC NO 1 OPEN CYCLE MHD-STEAM BOTTOMING NATURAL RESOURCE REQUIREMENTS

PARAMETRIC POINT	1	2	3	4	5	6	7	8
COAL, LB/KW-HR	.65889	.67189	.67048	.65889	.78213	.77504	1.03154	1.00859
SORBANT OR SEED,LB/KW-HR	.00232	.00274	.00258	.00232	.00318	.00305	.00308	.00317
TOTAL WATER, GAL/KW-HR	.592	.616	.621	.592	.664	.662	.660	.661
COOLING WATER	.578	.602	.607	.578	.555	.650	.651	.652
GASIFIER PROCESS H2O	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
CONDENSATE MAKE UP ,	.01041	.01059	.01089	.01041	.00791	.01125	.00811	.00817
WASTE HANDLING SLURRY	.0044	.0034	.0035	.0044	.0010	.0010	.0012	.0012
SCRUBBER WASTE WATER	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
NOX SUPPRESSION	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
TOTAL LAND ACRES/100MWE	52.98	60.24	69.05	52.98	53.97	53.69	54.99	54.88
MAIN PLANT	13.35	18.24	27.44	13.35	13.49	13.45	13.49	13.53
DISPOSAL LAND	13.79	14.07	14.03	13.79	11.27	11.17	12.32	12.05
LAND FOR ACCESS RR	25.84	27.94	27.58	25.84	29.21	29.07	29.17	29.30

PARAMETRIC POINT	9	10	11	12	13	14	15	16
COAL, LB/KW-HR	.65914	.65818	.69791	.65949	.71393	.59174	.63841	.62937
SORBANT OR SEED,LB/KW-HR	.00170	.00407	.00126	.00234	.00178	.00238	.00207	.00264
TOTAL WATER, GAL/KW-HR	.592	.590	.603	.562	.631	.506	.557	.543
COOLING WATER	.578	.576	.563	.549	.616	.494	.544	.530
GASIFIER PROCESS H2O	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
CONDENSATE MAKE UP ,	.01041	.01031	.01425	.01003	.01123	.00900	.00995	.00970
WASTE HANDLING SLURRY	.0049	.0033	.0063	.0033	.0037	.0030	.0033	.0032
SCRUBBER WASTE WATER	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
NOX SUPPRESSION	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
TOTAL LAND ACRES/100MWE	52.98	52.96	53.70	51.40	56.02	47.71	50.56	50.50
MAIN PLANT	13.35	13.34	13.41	13.40	13.58	13.29	13.31	13.33
DISPOSAL LAND	13.80	13.78	14.31	13.81	14.94	12.39	13.36	13.18
LAND FOR ACCESS RR	25.94	25.83	25.97	24.18	27.50	22.03	23.89	23.99

PARAMETRIC POINT	17	18	19	20	21	22	23	24
COAL, LB/KW-HR	.66838	.00000	.00000	.00000	.00000	.00000	.00000	.00000
SORBANT OR SEED,LB/KW-HR	.00235	.00000	.00000	.00000	.00000	.00000	.00000	.00000
TOTAL WATER, GAL/KW-HR	.611	.000	.000	.000	.000	.000	.000	.000
COOLING WATER	.596	.000	.000	.000	.000	.000	.000	.000
GASIFIER PROCESS H2O	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
CONDENSATE MAKE UP ,	.01042	.00000	.00000	.00000	.00000	.00000	.00000	.00000
WASTE HANDLING SLURRY	.0044	.0000	.0000	.0000	.0000	.0000	.0000	.0000
SCRUBBER WASTE WATER	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
NOX SUPPRESSION	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
TOTAL LAND ACRES/100MWE	53.66	.00	.00	.00	.00	.00	.00	.00
MAIN PLANT	13.46	.00	.00	.00	.00	.00	.00	.00
DISPOSAL LAND	13.99	.00	.00	.00	.00	.00	.00	.00
LAND FOR ACCESS RR	26.20	.00	.00	.00	.00	.00	.00	.00

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of capital cost combined with the reduction of speed treatment equipment costs produces energy costs which are slightly [∿ 0.556 mill/MJ (2 mills/kWh)] below the base case.

Points 7 and 8 use the lignite coal with 27 and 18% moisture respectively. Point 7 is an attempt to illustrate the problem associated with the very low-quality fuel gas. Both the gapor combustion air and the preheat stream are heated to 1525°K (2285°F) before being introduced to the separate air heater. Only a 40°K (72°F) increase in the temperature of the preheat stream is feasible. The MHD flame temperature of this case is below 2700°K (4400°F), which results in an increased magnet cost to offset the decrease in heat exchanger costs. For Point 8 the amount of MHD products circulated is very small, reducing the size of heat exchange equipment and resulting in energy costs below those of Points 5 and 6, since the overall efficiency is nearly identical.

Points 9, 10, and 11 have ash carry-overs from the combustor of 5, 20, and 100% respectively. The results are similar to those discussed under Base Case 2. Lower ash carry-over results in lower energy cost, but the case with 100% carry-over has the smallest seed makeup requirement.

No exhaust gas was recirculated for Point 12, and the combustor temperature went to 2953°K (4855°F) at a pressure of 1.2159 MPa (12 atm). Because of the high combustor temperature, the duct length was shorter (∿ 20%) than the base case at nearly twice the pressure ratio. This resulted in substantial savings in the superconducting magnet. The lower volumetric flow rate of the preheat stream and the lower intermediate temperature levels resulted in a very large reduction of heat exchanger costs. The direct capital cost of this point are about 20% lower than it is for the base case. Surprisingly, the efficiency is identical to that of the base case. This may be a result of a nonoptimized set of base case parameters (pressure ratio, generator loading coefficient, velocity coefficient, etc.). The total energy cost is more then 1.389 mills/MJ (5 mills/kWh) below that of the base case.

Point 13 has a lower preheat temperature 1668°K (2543°F) than the base case. It is very similar to the points which used the other coals. The heat exchanger costs are reduced substantially because of reduced temperature level and reduced recirculated flow. Due to the increase in stream power relative to MHD power, however, the efficiency is also down. The energy cost is about 0.278 mill/MJ (1 mill/kWh) below that of the base case.

Point 14 has a higher preheat temperature 2218°K (3532°F) than does the base case. The efficiency is increased substantially since much more of the power is generated by the MHD plants. The capital and O&M costs also increase, and the energy cost is slightly below that of the base case.

Points 15 and 16 have MHD pressure ratios of 8 and 10 respectively. In each case the efficiency and capital cost increases. In Point 15, however, the O&M costs are about 0.0167 mill/MJ (0.06 mill/kWh) below those of the base case so that its energy cost is about 0.0278 mill/MJ (0.1 mill/kWh) below that of the base case cost. For Point 16 the energy cost is 0.4167 mill/MJ (1.5 mills/kWh) above that of the base cases. The optimum pressure ratio probably lies between 6 and 8 for the standard economic conditions.

In Point 17 the use of a 16.6 MPa (2400 psig) steam plant results in decreased cost, power output, and efficiency. The energy cost ia about 0.0278 mill/MJ (0.1 mill/kWh) above that of the base case.

9.4.4 Natural Resource Requirements

The overall economic program also calculated the natural resource requirements for all parametric points. The results are summarized in Tables 9.15, 9.21, and 9.27 for Base Case Numbers 2, 3, and 1 and their variations, respectively.

The fuel requirements of the plants (per kWh) are inversely proportional to overall plant efficiency and should be lower for open-cycle MHD than for other cycles. For Base Case Numbers 1 and 2 there is no sorbent, and the seed requirements are related to the ash carry-over from the combustor and the EPA particulate emission standards in a complex manner. For ash carry-over between 5 and 20%, the ash collected is reinjected into the combustor. The seed which is lost in the

slag tap from the combustor increases with increasing carry-over and overall makeup increases. Leaching to recover this relatively small amount of seed is not justified. When 100% ash carry-over exists, however, it is necessary to leach the collected ash, and nearly all of the seed can be recovered. Hence, the makeup is actually reduced with 100% ash carry-over.

For Base Case Number 3, very large quantities of sorbent are required for sulfur treatment in the gasifier. The need to dispose of the spent sorbent results in a large disposal land requirement. The total land required for Base Case Number 3 is 80% greater than for Base Case Numbers 1 and 2.

The cooling water requirements are also nearly inversely proportional to efficiency, and should provide additional incentives for developing open-cycle MHD.

9.5 Capital and Installation Costs of Plant Components

The major components required for an open-cycle MHD power plant range from standard equipment items (for example, steam turbines which will be made up of standard building blocks) through items whose technology is well understood but which will require some design (for example, the main air compressors will be very similar to the compressor end of an industrial gas turbine) to components whose design can be charitably described as speculative (for example, heat exchangers to transfer the energy from the exhaust of the MHD duct to the stream to be preheated). Estimates of prices of such disparate items have to be made on different bases.

The list prices of standard equipment for the base cases (steam turbines and feedwater heaters) were obtained from the appropriate Westinghouse divisions. By plotting the data supplied for open-cycle MHD and for the nonequilibrium MHD base cases, price estimates for the open-cycle parametric points were made. This was possible because the same steam conditions, condenser pressure, and feedwater heating were used for both concepts.

For the air compressors, the Westinghouse Gas Turbine Division provided a breakdown of the price of industrial gas turbine components. This proprietary information was then used to estimate the price of the air compressor section.

The remaining major components are nonstandard items. The designs of these components are of a preliminary nature and there is a wide range of uncertainty about their validity. In some cases, the design is based on scaling of a reasonably well-understood technology (for example, the superconducting magnet); in the others, the design is a proposed means of dealing with a problem (for example, the recovery heat exchangers). While a correspondingly high degree of uncertainty must exist in price estimates of all these nonstandard items, those for the most speculative designs are most suspect. The designs and price estimates of these nonstandard items are discussed in Appendices A 9.1 through A 9.6 and A 9.9 through A 9.11.

9.5.1 Major Components for Three Base Cases

Tables 9.28, 9.29, and 9.30 contain the sizes, weights, and costs of the major components for open-cycle MHD Base Case Numbers 1,2, and 3, respectively. The air preheater is the largest single item for Base Case Numbers 1 and 2. Since both cases include the use of silicon carbide and high-nickel alloy heat recovery exchangers, both air preheaters also represent major areas of uncertainty. Although the technical uncertainty of the combustor and duct is probably as high as that of the air preheaters, their impact on the economics is much less, permitting much greater latitude in their design. For this reason, the air preheater appears to be the critical item in the development of an economically successful open-cycle MHD power plant.

The magnet and inverter system also represent large fractions of the plant costs. The technology of these appears to be well in hand, but the cost estimates must be considered as uncertain. This is particularly true of the magnet whose price is based on a wire price and current density which are not currently attainable. If these reduced prices

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Table 9.28 Sizes, Weights and Prices of Major Components for Base Case 1

Major Component	Size,ft L	W	H	Weight,lb	Cost FOB Mfg. Plant,$	Units Req'd	Total Cost,$
Combustor	50	50	80	1,306,000	1,413,000	1	1,413,000
Generator Duct	66	9	9	396,000	982,000	1	982,000
Magnet	66	30ϕ		3,400,000	64,000,000	1	64,000,000
Air Preheater	800	200	100	98,300,000	122,000,000	1	122,000,000
Seed Recovery	400	300	150	33,000,000	40,500,000	1	40,500,000
Inverter	600	200	30	9,000,000	57,200,000	1	57,200,000

ϕ = diameter

Table 9.29 Sizes, Weights and Costs of Major Components for Base Case 2

Major Component	Size,ft L	W	H	Weight,lb	Cost FOB Mfg. Plant,$	Units Req'd	Total Cost,$
Combustor	50	50	80	888,000	643,000	1	643,000
Generator Duct	72	9	9	280,000	652,000	1	652,000
Magnet	72	30ϕ		3,680,000	69,000,000	1	69,000,000
Air Preheater	500	50	50	19,400,000	93,000,000	1	93,000,000
Seed Recovery	400	300	150	36,000,000	44,500,000	1	44,500,000
Inverter	600	200	30	9,300,000	57,800,000	1	57,800,000

ϕ = diameter

Table 9.30 Size, Weights and Costs of Major Components for Base Case 3

Major Component	Size, ft L	W	H	Weight, lb	Cost FOB Mfg. Plant, $	Units Req.d	Total Cost, $
Combustor	50	30ϕ		705,000	691,000	1	691,000
Generator Duct	177	8	8	674,000	1,584,000	1	1,584,000
Magnet	177	26ϕ		8,000,000	163,000,000	1	163,000,000
Air Preheater	500	50	50	10,600,000	49,000,000	1	49,000,000
Seed Recovery	200	200	150	20,000,000	21,800,000	1	21,800,000
Inverter	600	250	30	10,500,000	66,400,000	1	66,400,000

ϕ = diameter

and increased densities do not materialize, the magnet price could escalate sharply.

The balance of plant equipment costs is calculated in the overall economic program according to algorithms developed by other contributors to this program. Such items as coal crushers, fuel oil shortage, gasifiers, and carbonizers are included in this category.

9.5.2 Balance of Plant

The price of the seed treatment plant is somewhat uncertain because of a lack of fundamental engineering data. The estimates presented, however, are based on conservative judgments.

Furthermore, half of the seed recovery costs of Base Case Numbers 1 and 2 are due to the collection device (electrostatic precipitator), so that the total seed recovery costs should not be a major source of uncertainty.

9.6 Analysis of Overall Cost of Electricity

The results presented in Section 9.4 indicate that the energy costs for the open-cycle MHD plants range from 7.5 to 9.72 mills/MJ (27 to 35 mills/KWh) (Tables 9.14, 9.20 and 9.26). Of these costs, more than 70% are capital charges and about 20% are fuel costs. By conventional standards this split is rather heavy on the capital side for a base-load plant, and some justification is in order. The capital costs are given for plants started in 1974 which had construction times of 7 to 8 years; they include escalation and so are expressed in 1981 and 1982 dollars. For such a plant, the levelized fuel costs should correspond to fuel prices in 1989 and 1990. It is not difficult to imagine a doubling or perhaps tripling of the level used for Tables 9.14, 9.20 and 9.26 [\$0.806/GJ (\$0.85/10^6 Btu)] in this span. If both the O&M and fuel charges were to double, the capital charges would become a more normal fraction of total energy costs (< 66%); if they tripled, the capital charges would be less than 55% of the total. In this study, a capacity factor of 0.65 was used, and this (compared to normal study values of 0.8) also contributes to the dominance of capital charges.

In evaluating the energy costs of these cycles, one should keep in mind the considerable uncertainty in the price estimates. This is especially important when capital represents such a large part of total energy costs.

The primary uncertainty is in the recovery heat exchangers. As mentioned in Section 9.1, there is no established technology for these components, and prices must necessarily be uncertain. Although there are other items of uncertain technology (for example, generator duct and combustors), they represent rather small parts of the overall plant costs. The direct cost of the recovery heat exchanger is estimated to be over \$60/kW or about 20% of the direct equipment costs for the direct-fired cycle (Base Case Number 2).

Another area of uncertainty is the coupling heat exchanger (steam generator). Since it operates at much lower temperatures, it does not appear to present as severe a problem as the recovery exchanger. The existence of the slag-seed mixture in both liquid and solid states, however, does present corrosion and fouling problems (Reference 9.4). The direct cost of the steam generator is about \$25/kW, or more than 8% of the Base Case Number 2 direct cost.

Although the technology of the superconducting magnet is apparently well understood, its price must be considered highly uncertain at this time. The estimates given here are based on superconductor costs which are much lower than presently exist and which appear to be optimistic. The structural design used does not provide for restraining the cross-over wires, and no consideration has been given to transient forces which might occur during a load-trip. As estimated, the direct cost of the magnet is over \$35/kW or 12% of the Base Case Number 2 direct cost.

In summary, although the energy cost provides a valuable guide in comparing cases, the limitations must be kept in mind at all times.

9.7 Conclusions and Recommendations

9.7.1 Conclusions

- If the technical problems can be solved economically, open-cycle MHD plants can achieve high overall efficiencies with a wide variety of coals.
- The technology of heat recovery apparatus and superconducting magnets is not sufficiently advanced to permit accurate estimates of direct capital costs.
- With high-sulfur coal, the use of the seed material to remove sulfur from the exhaust products requires large power and heat inputs.

9.7.2 Recommendations

- Future efforts on open-cycle MHD should focus on
  - Design and cost of recovery heat exchangers
  - Design and cost of coupling heat exchangers
  - Cost projections for very large superconducting magnets.
- The final choice of the cycle to be used should be delayed until the recovery heat exchanger solution is better defined.
- For high-sulfur coals, the use of conventional sulfur removal techniques should be considered. The potassium seed could be recycled in the sulfate form, and excess sulfur could be removed from the stack gases by conventional techniques.

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Appendix A 9.1

OPEN-CYCLE MHD SEED TREATMENT

A 9.1.1 General Discussion

The economic feasibility of open-cycle MHD plants depends on recovering and returning to the process a high percentage of the seed used in the MHD duct. Currently, it is estimated that at least 95% (Reference 9.5) recovery is required to maintain economic feasibility. In two of the three base cases, potassium carbonate seed is utilized to remove the sulfur found in the coal as well as to provide the required conductivity in the duct. Unless potassium carbonate is regenerated from the potassium sulfate which is removed by the electrostatic precipitator at the end of the system, sufficient makeup potassium carbonate must be added to remove the sulfur while discarding most of the potassium sulfate. This procedure does not allow the 95% recovery criterion to be met, even for the two low-sulfur coals studied. In Base Case 3, cesium carbonate (in the form of pollucite ore) is used to provide conductivity in the duct. No regeneration is necessary since, in this case, the sulfur is removed in the gasifier, prior to the combustion of the fuel in the MHD combustor.

The regeneration process evaluated is based upon work carried out by the U.S. Bureau of Mines (Reference 9.6). Their work showed that the reaction

$$4H_2 + K_2SO_4 \rightarrow 2KOH + H_2S + 2H_2O \qquad \text{(A 9.1.1)}$$

takes place at elevated temperatures in pure hydrogen. Their findings were that the rate of this reaction reached a maximum at about 1048°K (1427°F).

Two methods considered for producing the needed hydrogen were:

- The electrolysis of water
- Gasification of coal using a Westinghouse fluidized bed gasifier.

The first method, electrolysis, presented the advantages of simplicity in design and use, and the use of pure hydrogen (for which kinetic data were available). There were, however, two major disadvantages. First, the energy efficiency of an electrolysis unit is very low; and, therefore, the energy required to provide the same amount of hydrogen as a gasifier was a factor of two higher. Second, the cost of an electrolysis unit is about two and one-half times greater than an oxygen-blown gasifier producing a like amount of reducing agent. See Subappendix AA 9.1.1.

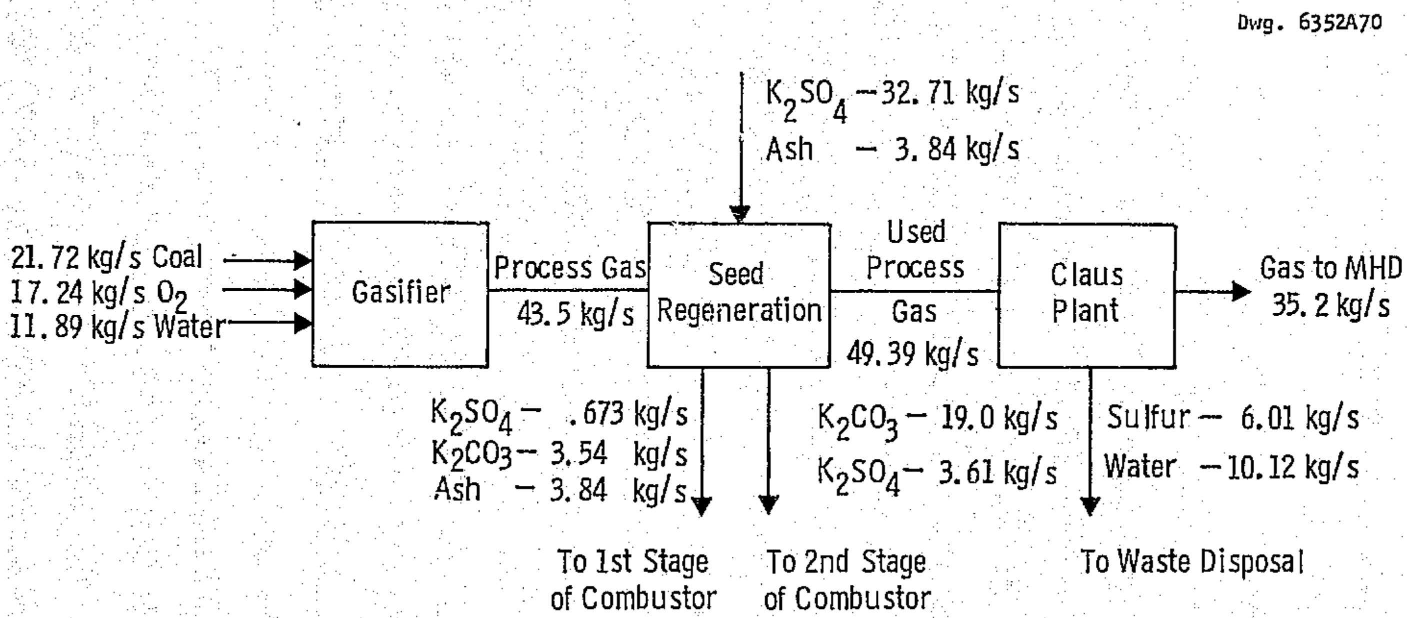
Based on the above considerations, the use of a coal gasifier was chosen. The coal gasifier produces a gas stream containing hydrogen sulfide, carbon monoxide, carbon dioxide, hydrogen, water vapor, and nitrogen (if an airblown gasifier is used). In addition to the above reaction, the following reaction also takes place

$$CO_2 + 2KOH \rightarrow K_2CO_3 + H_2O. \qquad \text{(A 9.1.2)}$$

In addition to these two reactions of the synthesis gas with the potassium seed, the water gas shift reaction

$$CO + H_2O \rightleftarrows CO_2 + H_2 \qquad \text{(A 9.1.3)}$$

tends to maintain a fixed equilibrium between the respective gases. Since one of the above reactions uses hydrogen and another carbon dioxide, and since water is produced by both gas-solid reactions, the carbon monoxide present in the synthesis gas will continually be shifted to hydrogen. This being the case, the carbon monoxide contained in the synthesis gas can be counted in determining the hydrogen available for reaction. This process is shown conceptually in Figure 9.1.1 (based on Base Case 2).



Stream Name	Process Gas	Used Process Gas	Gas to MHD
Temperature, °K	1144	1040	899
Pressure, KPa	1520	1499	608
H_2 Mole Fract.	0.2788	0.1522	0.2167
CO	0.3764	0.2055	0.2927
CO_2	0.1464	0.2753	0.3921
H_2O	0.1324	0.2078	0.0128
H_2S	0.0110	0.0990	0
CH_4	0.0551	0.0601	0.0856

Fig. A 9.1.1 — Simplified diagram of seed regeneration system based on Base Case 2

The use of a gasified coal as the reductant in regeneration of the seed presented two difficulties. First, kinetic data were not available for the gas-solid reactions. Second, the complexity of the process was increased. The first difficulty was overcome by noting that the gas-solid reactions were essentially irreversible (therefore, the presence of hydrogen sulfide in the synthesis gas would have little if any effect). Also, it was assumed that the addition of other gases to the hydrogen would only dilute it and not prevent it from reacting.

The next question to be answered was whether the use of an air-blown or oxygen-blown gasifier would be more advantageous. The overall efficiency and total system costs were not greatly affected by whether the gasifier was oxygen-blown or airblown. An oxygen-blown system, however, does appear to have a major advantage over an airblown one: the gas not used in the regeneration system and returned to the MHD plant has a heating value almost twice as large as the gas from an airblown system. Also, since the mass flow is about half as large for an oxygen-blown system and since both systems are saturated with water at 332°K (138°F), there is a lower mass flow of water back to the MHD plant with an oxygen system. Based on the above observations, an oxygen-blown gasification system was chosen.

A 9.1.2 Regeneration Process Description

Figure A 9.1.2 presents a simplified flow diagram for the process. A flow chart is presented in Table A 9.1.1. The oxygen plant product (flow 4 in Figure A 9.1.2) is essentially pure oxygen gas at 300°K (80.31°F) and 0.1013 MPa (1 atm). This stream is then compressed to 1.722 MPa (17 atm) in the compression unit and the temperature raised to 699°K (799°F). A small part of the compressed oxygen stream is fed to the sulfur dioxide burner in the Claus plant, while the remainder is sent to the coal gasification system.

The gasifier includes a series of coal-handling and preparation steps, along with the actual volatilization and gasification of the coal. Steam at 478°K (400°F) and 1.722 MPa (17 atm), coal, and the flow 4 oxygen

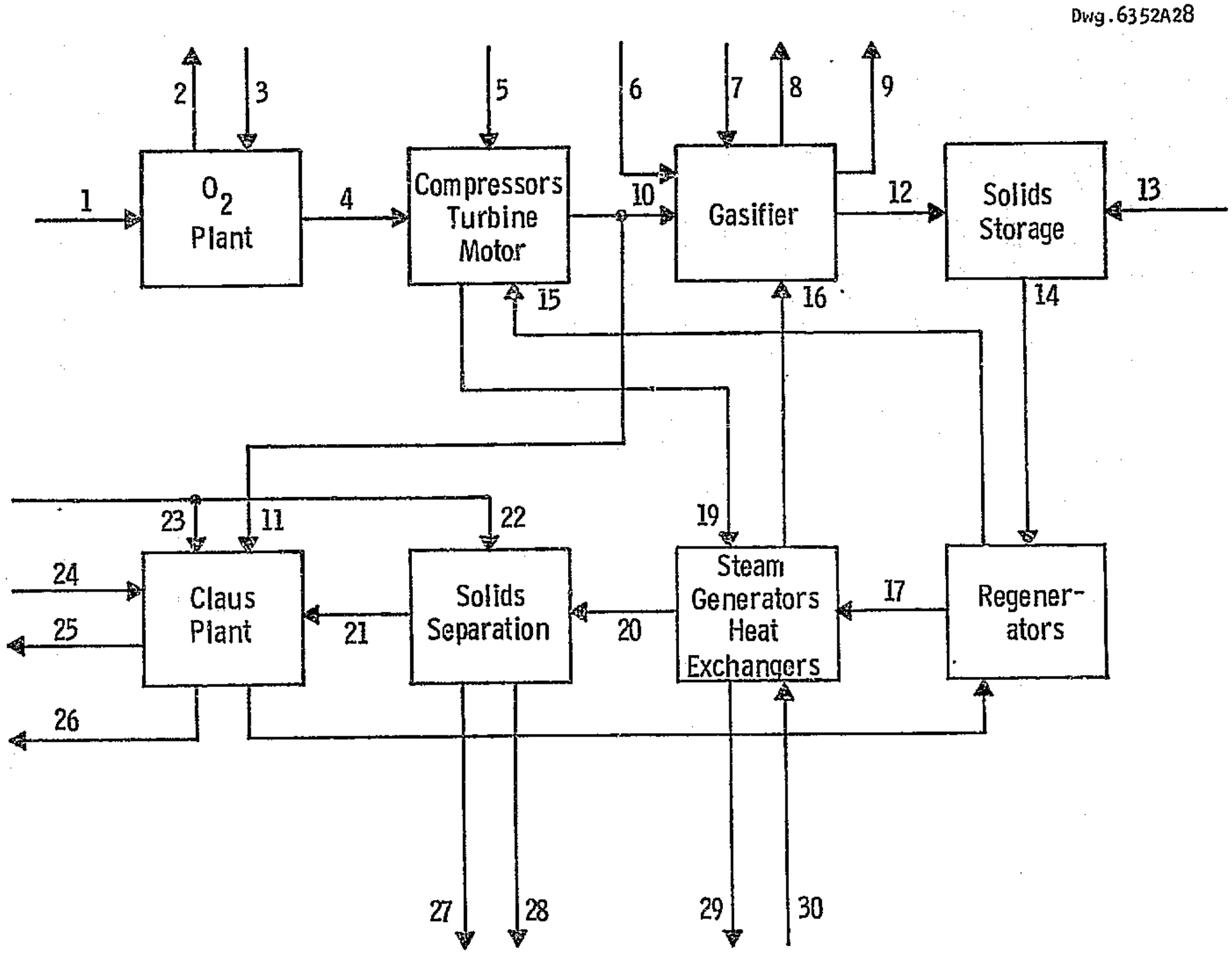


Fig. A 9.1.2—Simplified flow diagram of seed regenerative system for Base Case 2

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TABLE A 9.1—FLOW CHART FOR SEED REGENERATION SYSTEM FOR BASE CASE 2

Flow Name	Air Feed	N_2 Rejection	Power	O_2	Power	Coal Feed	Power	Ash Reject	Exhaust	O_2	O_2	Process Gas	Seed + Ash	Solids + Gas	Warmed Gas
Flow Number	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
Mass Rate, kg/s	73.99	56.75		17.24		21.72		2.09	7.18	14.2	3.04	43.5	36.55	80.05	36.3
Temperature, °K	300	300		300		300		1144	370	669	669	1144	350	960	760
Pressure, kPa	101	101		101		101		1520	101	1722	1722	1520	101	1520	1418
Power, kW			16136		1483		331.65		0.7066						
N_2, Mole Fraction	0.79	1.0		0					0	0	0	0		0	0
H_2	0	0		0					0	0	0	0.2788		0.2788	0.2167
CO	0	0		0					0.1632	0	0	0.3764		0.3764	0.2927
CO_2	0	0		0					0.1048	0	0	0.1464		0.1464	0.3921
H_2O	0	0		0					0	0	0	0.1324		0.1324	0.0128
H_2S	0	0		0					0	0	0	0 .011		0 .011	0
CH_4	0	0		0					0	0	0	0.0551		0.0551	0.0856
SO_2	0	0		0					0.0254	0	0	0		0	0
O_2	0.21	0		1.0					0	1.0	1.0	0		0	0
S, kg/s													0	0	
K_2SO_4													32.71	32.71	
K_2CO_3													0	0	
Ash								2.09					3.84	3.84	
Molecular Weight (Gas)	28.6	28							29.66	32	32	21.3		21.3	27.5

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TABLE A 9.1 — FLOW CHART FOR SEED REGENERATION SYSTEM FOR BASE CASE 2 (cont'd)

Flow Name	Steam	Process Gas	Cold Gas	Expanded Gas	Cool Process Gas	Cool Process Gas	Power	Power	Cool Water Make	Evap. Water	Sulfur + Water	K_2CO_3 K_2SO_4	Ash + Seed	Hot Gas to MHD	Steam Water
Flow Number	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30
Mass Rate, kg/s	11.89	80.05	36.3	36.3	80.05	49.39			8.6	8.6	16.13	22.61	8.05	35.2	11.89
Temperature, °K	478	1040	332	598	541	541			300	373	420	541	541	899	478
Pressure, kPa	1722	1499	1428	607	1469	1444			101	101	1418	1418	1418	607	1722
Power, kW							9.2	221							
N_2, Mole Fraction	0	0	0	0	0	0			0	0	0			0	0
H_2	0	0.1522	0.2167	0.2167	0.1522	0.1522			0	0	0			0.2167	0
CO	0	0.2055	0.2927	0.2927	0.2055	0.2055			0	0	0			0.2927	0
CO_2	0	0.2753	0.3921	0.3921	0.2753	0.2753			0	0	0			0.3921	0
H_2O	1.0	0.2078	0.0128	0.0128	0.2078	0.2078			1.0	1.0	1.0			0.0128	1.0
H_2S	0	0.0990	0	0	0.0990	0.0990			0	0	0			0	0
CH_4	0	0.0601	0.0856	0.0856	0.0601	0.0601			0	0	0			0.0856	0
SO_2	0	0	0	0	0	0			0	0	0			0	0
O_2	0	0	0	0	0	0			0	0	0			0	0
S, kg/s		0									6.01	0	0		
K_2SO_4		4.28									0	3.61	0.673		
K_2CO_3		22.54									0	19.0	3.54		
Ash		3.84									0	0	3.84		
Molecular Weight (Gas)	18	26.3	27.5	27.5	26.3	26.3			18	18	18			27.5	18

stream were used to make a synthesis gas at 1.520 MPa (15 atm) and 1144°K (1600°F) (flow 12). This hot synthesis gas entrains solids mixed with it from the solids storage system.

The solids are a mixture of ash and potassium sulfate obtained from an electrostatic precipitator on the stack of the MHD plant (flow 13).

The gas-solid mixture (flow 14) is then sent to the regenerators. The enthalpy of the synthesis gas stream is used to heat the product gas-solids flow up to near the required reaction temperature [960°K (1268°F)]. Since the net reactions are exothermic, some cooling is required to keep the regenerator temperature below 1048°K (1427°F). This is accomplished by using cooling gas to cool internal heat exchangers in the regenerator. Figure 9.1.3 presents a sketch of this regenerator. The exact location of the various heat exchanger modules must be determined by a detailed modeling study of this reactor. The detailed design specifications are presented in Subappendix 9.1.3. The cooling gas comes from the Claus plant (flow 18) at 332°K (138°F) and is then sent, after being heated in the regenerators, to the turbines to be expanded from 1.418 to 0.607 MPa (14 to 6 atm), and cooled from 760°K to 598°K (908 to 616°F) (flows 15 and 19).

The product gas-solids mixture (flow 17) is then sent to heat exchangers and steam generators to be cooled to 541°K (514°F). Heat exchangers are first used to lower the temperature to around 854°K (1078°F). The cooling gas (flow 19) comes from the expanders at around 598°K (617°F). This flow is heated and sent to the MHD plant (flow 29) after a small amount is bled off to the coal drying unit in the coal gasification unit. The product gas-solids flow is then sent to a set of steam generators to be cooled to a final temperature of 541°K (514°F) and sent to solids separation (flow 20). The steam produced is sent directly to the gasifier system (flow 16).

The solids separation system consists of a cyclone to remove the ash (and solids intimately mixed with the ash) and an electrostatic precipitator to remove the remainder of the solids (mainly seed materials).

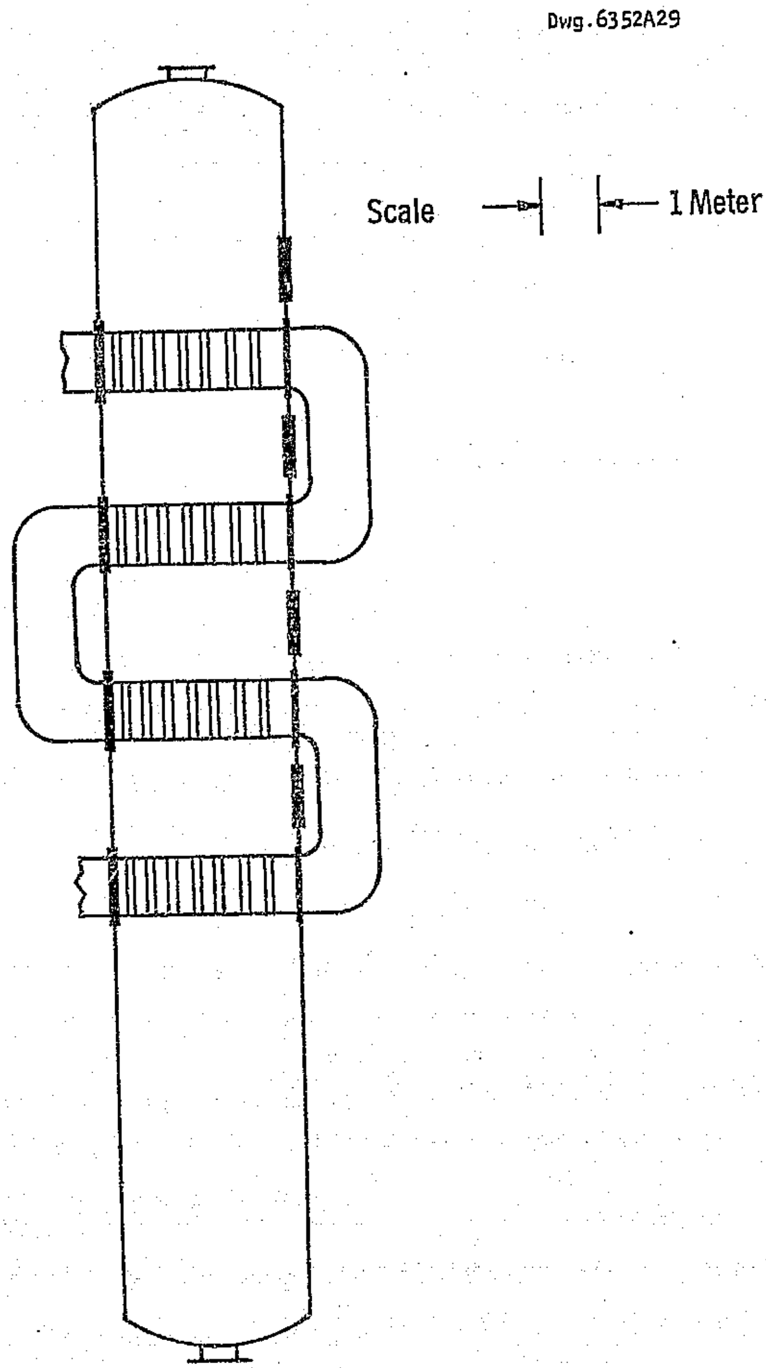

Fig. A 9.1.3—Regenerator configuration for Base Case 2

This separation is necessary since ash is removed only from the first stage of the combustor. To prevent a build-up of ash in the system, the ash must be separated from the seed and recycled to the first stage of the combustor. It is not desirable to return the seed and ash to the first stage, since 20% (Reference 9.7) of the seed introduced into the first stage is lost to the slag; but only 2% is lost to the slag if the seed is introduced in the second stage (Reference 9.3). To minimize the amount of seed loss from the system therefore, it is desirable to achieve the maximum separation possible between ash and seed. The basis for the separation system utilized is the work by Heywood and others (Reference 9.4). Their work indicated that the average particle size of ash was about 10 μm, while the average seed diameter is about 1.2 μm. A relatively clean separation can be made, under these conditions, of the ash from the seed by the use of a cyclone. A cyclone, however, was not practical for use to remove seed from the gas stream because of the small size of the particles. An electrostatic precipitator was utilized, enclosed in a pressure vessel. The precipitator is sized to remove 99.5% of the solids from the gas stream. The two solids streams are then sent to their respective sections of the combustor.

The process gas stream, with the solids removed (flow 21), is sent to a Claus plant. This plant contains a sulfur dioxide generator where sulfur from the Claus reactor is burned with oxygen; a Claus reactor where the sulfur dioxide and hydrogen sulfide are mixed and react as

$$SO_2 + 2H_2S \rightarrow 3S + 2H_2O; \qquad \text{(A 9.1.4)}$$

a scrubber/demister where the sulfur (entering as a vapor) is condensed, solidified, and washed from the gas stream; a cooling tower to provide cooling water to the scrubber/demister; and a settling tank to allow the solid sulfur to settle out of the scrubber water. The gas leaves this unit cooled to 332°K (138°F) and saturated with water (flow 18).

The flow chart for this regeneration system, Figure A 9.1.2, is based on a system using Illinois No. 6 coal and 20% carry-over of the ash from the combustor.

A 9.1.3 Overall Potassium Balance

In order to size the seed regeneration plant, an overall potassium balance must be performed on the MHD system and seed regeneration system. The derivation of the pertinent equations is presented in Subappendix AA 9.1.1 for Base Case 2 (Illinois No. 6 coal, as received, and 20% ash carry-over in the combustor). Equations A 9.1.5 through A 9.1.10 were derived where

Y_I = mass fraction of ash carried over from the combustor

Y_s = fraction of sulfur to be removed from the coal

Y_p = fraction of particulates to be removed from the coal

Y_c = fraction of potassium sulfate converted to potassium carbonate.

and the S's are defined in Figure AA 9.1.2.1 in Subappendix AA 9.1.2. These equations are iterated for various values of Y_c until S_3 in Equations A 9.1.10 and A 9.1.8 are equal.

Using these equations, the percent regeneration was found to be 86.9%, the ash rate into the regeneration system 3.84 kg/s, and the potassium sulfate rate into the regeneration system 32.71 kg/s for Base Case 2 (Illinois No. 6 coal, 20% ash carry-over, and 3% moisture). The amount of makeup potassium carbonate needed was 1.24 kg/s and the electrostatic precipitator capture efficiency required was 99.53%.

Based on these values, a seed regeneration system for Base Case Case 2 was designed. Subappendix AA 9.1.3 contains the detailed calculations. The cost for this system was found to be $29,510,000, and it used 9.07 kg/s of coal and required 18.181 MW of electrical power. Of the total coal rate needed to run the MHD plant, 6.19% is required to operate the seed regeneration system. The mechanical equipment group includes the coal reclaimer conveyor, the coal crusher/dryer, the predried coal elevator, the ash slurry pumps, and the electric motor. The heat exchange equipment group includes the heat exchangers and steam generators. The rotating machinery group includes the lock gas compressor, the oxygen compressor, and the turbines. The vessels group includes both pressurized and unpressurized vessels. Areas of greatest cost are the oxygen plant and vessels.

$$Y_e = \frac{(1 - Y_p)}{\frac{Y_I\ (1 + 0.26\ (1 - Y_e))}{(1 - (1 - Y_e)\ Y_I)} + \frac{0.0229\ (S_2 + S_1)}{X_{ASH,1}\ S_1} - 0.0229 - \frac{0.145\ X_{K,1}}{X_{ASH,1}}} \tag{A 9.1.5}$$

where Y_e is the fraction of particulates allowed to be discharged up the stack.

$$X_{ASH,5}\ S_5 = \frac{(1 - Y_E)\ X_{ASH,1}\ S_1\ Y_I}{(1 - (1 - Y_E)\ Y_I)} \tag{A 9.1.6}$$

where $X_{ASH,5}\ S_5$ is the ash flow in the regeneration system.

$$X_{K,8}\ S_8 = \frac{(1 - Y_E)\ X_{ASH,1}\ S_1\ Y_I}{(1 - (1 - Y_E)\ Y_I)} \left(\frac{0.47\ Y_c - 0.410}{0.958\ (1 - Y_c)}\right) + \frac{0.01\ S_2 + S_1\ (0.01\ (1 - X_{ASH,1}) + 0.34\ X_{K,1} - 2.4\ X_{S,1}\ Y_s)}{0.958\ (1 - Y_c)} \tag{A 9.1.7}$$

where $X_{K,8}\ S_8$ is the potassium flow from the regeneration system to the second stage of the combustor.

$$S_3 = S_1\ (2.49\ X_{S,1}\ Y_s - 0.714\ X_{K,1}) - Y_c\ S_8\ X_{K,8} - \frac{(1 - Y_E)\ X_{ASH,1}\ S_1\ Y_I}{(1 - (1 - Y_E)\ Y_I)}\ (0.061 + 0.49\ Y_c) \tag{A 9.1.8}$$

where S_3 is the potassium flow in the makeup stream.

$$X_{K,5}\ S_5 = 0.6\ X_{ASH,5}\ S_5 \tag{A 9.1.9}$$

where $X_{K,5}\ S_5$ is the potassium flow from the regeneration system to the first stage of the combustor.

A potassium balance around the system is given in Equation A 9.1.10.

$$S_3 = (0.02 + 0.98\ Y_e)\ (S_3 + X_{K,8}\ S_8) + X_{ASH,5}\ S_5 (0.12 + 0.48\ Y_e) - 0.8\ X_{K,1}\ S_1 \tag{A 9.1.10}$$

Table A 9.1.2 - Equipment Costs for Seed Regeneration System for Base Case 2

Equipment Type	Cost
Rotating Machinery	$ 980,000
Heat Exchange Equipment	$ 1,160,000
Vessels	$10,760,000
Mechanical Equipment	$ 1,560,000
Oxygen Plant	$15,050,000

System	Cost
Oxygen Plant	$15,050,000
Compressors-Turbine-Motor	$ 1,070,000
Gasifier	$ 4,990,000
Solids Storage	$ 600,000
Regenerator	$ 2,980,000
Steam Generators and Heat Exchangers	$ 1,160,000
Solids Separation	$ 2,440,000
Claus Plant	$ 1,220,000

In the cases where 100% ash carry-over is specified in the combustor, it will be necessary to install a leaching plant along with the seed regeneration plant. The inlet to the leaching plant is the ash-seed mixture obtained from the cyclone. The ash inlet is mixed with 373°K (212°F) water at 1.418 MPa (14 atm). Sufficient water is added to dissolve the seed. The ash is removed from the bottom of the settling tank as a slurry. The saturated liquid is then flashed. The liquid-seed slurry is fed to a scraped-film evaporator where the pure seed is formed and returned to the second stage of the combustor. The steam from the evaporator and flash drum is condensed to saturated water at 373°K (212°F) and returned to the leaching tank. Subappendix AA 9.1.4 presents the details in the design of this system.

In order to use the material balances presented in Subappendix AA 9.1.2 for Base Case 1, effective coal rate, ash content, and sulfur content must be computed. Subappendix AA 9.1.5 presents the details of the changes required.

It was found that five factors affect the cost and energy requirements of the seed regeneration system. These are: 1) flow rate of ash into the system; 2) flow rate of potassium sulfate into the system; 3) the percent regeneration of the potassium sulfate required; 4) the coal factor (a factor used to correct for the difference in heating value and composition of the various kinds of coal); and 5) the MHD gas factor (same reason as for the coal factor). The vessels whose costs are determined by the ash flow rate are the ash cyclone lockhoppers. The vessel cost determined by the flow rate of potassium sulfate is the potassium carbonate lockhopper. The vessels whose costs are determined by the combined flow of the ash and potassium sulfate are the potassium sulfate surge bins and potassium sulfate locks. The remaining equipment costs are a function of the product of potassium sulfate flow rate and the fractional conversion. These pieces of equipment either handle the process gas or are instrumental in producing the process gas. Since the amount of process gas is a direct function of the amount of sulfur to be removed, these vessel sizes and costs are functions of the product of the potassium

sulfate flow rate and the percent regeneration. Using the costs developed for Base Case 2, the energy and cost requirements were found as functions of the above variables. Details of these calculations are presented in Subappendix AA 9.1.6. The cost of the regeneration and leaching systems were broken down to the factors given in Equations A 9.1.11 through A 9.1.16 where $K_2\ SO_4$ and Ash as used in these equations represent the mass flow rate of the potassium sulfate and ash in kilograms per second. The fractional conversion of the potassium sulfate to potassium carbonate is represented by F_K.

Cost of Vessels, $ =

$$966509\ [(K_2SO_4)(F_K)]^{0.666} + 54492\ (K_2SO_4 + Ash)^{0.666} + 130507\ (Ash)^{0.666} + 82147\ (K_2SO_4)^{0.666} \qquad \text{(A 9.1.11)}$$

Cost of Rotating Machinery, $ =

$$120735\ [(K_2SO_4)(F_K)]^{0.59} + 13161\ (K_2SO_4 + Ash)^{0.59} \qquad \text{(A 9.1.12)}$$

Cost of Mechanical Equipment, $ =

$$54881\ [K_2SO_4)(F_K)] \qquad \text{(A 9.1.13)}$$

Cost of Heat Exchange Equipment, $ =

$$107729\ [(K_2SO_4)(F_K)]^{0.71} \qquad \text{(A 9.1.14)}$$

Cost of Oxygen Plant, $ =

$$962000\ [(K_2SO_4)(F_K)]^{0.8215} \qquad \text{(A 9.1.15)}$$

Cost of Leaching System, \$ =

$$99161 \{1 + 0.0227 (F_K)(Ash) + 0.45 [(F_K)(Ash) (2.23 - \frac{73.3 - K_F}{97.5 - F_K})^{0.55}] + 0.175 [(Ash)(F_K)]^{0.65}\} \quad \text{(A 9.1.16)}$$

The energy requirements, as a function of the above variables, are given by Equations A 9.1.17 through A 9.1.20.

Electric Power Required for Regeneration, kW =

$$0.612 (F_K) (K_2SO_4) \quad \text{(A 9.1.17)}$$

Coal Required for Regeneration, kg/s =

$$0.746 (F_K) (K_2SO_4) (CF) \quad \text{(A 9.1.18)}$$

where CF is the coal factor equal to 1, 0.925 and 1.03 for the bituminous, subbituminous, and lignite coals, respectively.

Steam Required for Leaching, kg/s =

$$2.13 (Ash) (F_K) [1 - \frac{0.446 (73.3 - F_K)}{(97.6 - F_K)}] \quad \text{(A 9.1.19)}$$

Electric Power Required for Leaching, kW =

$$2 + 4.58 (Ash) (F_K) \quad \text{(A 9.1.20)}$$

The energy available to the MHD cycle from the seed regeneration process is given by Equation 9.1.21. This energy is available at a temperature of 899°K (1159°F).

Energy Available to MHD Gas from the Regeneration, kW =

$$11147 (K_2SO_4) (F_K) (MHDF) \quad \text{(A 9.1.21)}$$

where MHD factor, MHDF, has a value of 1, 0.6, and 0.413 for the bituminous, subbituminous, and lignite coals, respectively.

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TABLE A 9.1.3–BASE CASE 1 RESULTS

Parametric Point	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17
MHD Coal, kg/s	156.2	94.68	47.81	156.2	187.1	189.0	235.0	240.1	156.2	156.2	156.2	155.2	164.6	141.0	151.9	148.7	156.2
Regeneration Coal, kg/s	7.432	4.486	2.319	7.432	3.029	3.128	3.646	3.725	7.484	7.28	13.77	7.216	7.197	6.670	7.251	7.041	7.432
Regeneration Power, kW	14190	8562	4427	14190	4076	4209	4849	4954	14280	13900	23650	13770	15110	12730	13840	13440	14190
Regeneration Cost × 10^{-6}, $	29.54	16.76	10.28	24.50	10.52	10.72	11.70	11.90	24.58	24.33	37.11	23.69	25.58	22.78	24.12	23.62	24.54
Precipitator Eff., %	99.58	99.57	99.62	99.57	99.50	99.49	99.51	99.51	99.56	99.60	99.48	99.42	99.51	99.63	99.58	99.59	99.57
Make-up*, kg/s	.5748	.4042	.2175	.5748	.7740	.7442	.7507	.7862	.4228	1.011	.1424	.575	.4304	.5934	.5149	.6555	.5748
Percent Recovery	97.79	97.44	97.60	97.79	96.61	96.71	96.56	96.59	98.37	96.12	99.4	96.9	98.19	97.82	98.02	97.48	97.79
Percent Ash Carryover	10	10	10	10	10	10	10	10	5	20	100	10	10	10	10	10	10
Cost per kW $/kW	14.99	14.30	17.25	14.99	5.951	5.529	6.056	6.188	12.48	12.34	18.92	12.12	13.34	11.50	12.19	11.99	12.63
	10.69	10.80	11.01	10.69	3.54	3.60	3.34	3.43	10.76	10.45	20.52	10.44	10.72	9.53	10.38	10.61	10.85

*K_2CO_3

TABLE A 9.1.4–BASE CASE 2 RESULTS

Parametric Point	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17
MHD Coal, kg/s	160.7	97.12	50.22	158.7	175.4	188.2	245.7	237.0	160.7	158.7	160.7	160.7	154.9	154.9	152.5	160.7	153.7
Regeneration Coal, kg/s	8.796	5.315	2.773	8.716	3.172	3.443	4.523	4.370	8.933	8.772	8.993	16.55	8.472	8.508	8.343	8.793	8.424
Regeneration Power, kW	16790	10140	5293	16640	4267	4631	6015	5811	17050	16740	17160	28070	16167	16240	15920	167[illegible]0	16080
Regeneration Cost × 10^{-6}, $	27.83	19.00	11.62	27.64	11.13	11.72	13.93	13.59	27.97	27.59	28.01	42.15	27.54	27.13	26.73	27.82	26.93
Precipitator Eff., %	99.53	99.53	99.51	99.53	99.56	99.53	99.56	99.57	99.50	99.50	99.47	99.48	99.57	99.53	99.53	99.53	99.53
Make-up*	1.054	.6407	.2723	.9650	1.200	1.235	1.316	1.257	.5788	.6908	.3612	.0665	.9114	.9378	1.000	1.054	.9652
Percent Recovery	95.85	95.83	96.43	96.17	95.43	95.24	95.27	95.37	97.72	97.25	98.58	99.74	96.39	96.17	95.85	95.85	96.04
Percent Ash Carryover	20	20	20	20	20	20	20	20	10	10	5	100	20	20	20	20	20
Cost per kW	13.96	15.95	19.97	13.89	5.820	5.952	7.136	6.898	14.03	13.86	14.05	21.28	13.99	13.72	13.55	14.15	13.55
	12.59	12.69	12.94	12.55	3.81	4.04	4.31	4.18	12.80	12.62	12.89	24.76	12.68	12.29	12.06	12.77	12.14

*K_2CO_3

TABLE A 9.1.5–BASE CASE 3 RESULTS

Parametric Point	1	2	3	4	5
MHD Coal, kg/s	147.2	90.57	45.83	139.9	134.9
Regeneration Coal, kg/s	0	0	0	0	0
Regeneration Power, kW	0	0	0	0	0
Regeneration Cost × 10^{-6}, $	0	0	0	0	0
Precipitator Eff., %	99.53	99.52	99.53	99.53	99.53
Make-up*, kg/s	.8400	.5190	.2647	.8045	.7785
Percent Recovery	97.47	97.47	97.47	97.47	97.47

*Pollucite

The total cost is determined by adding together all the applicable factors. The net energy rate is determined by subtracting the energy available to MHD gas from the sum of the product of the coal rate, the higher heating value of the coal, the product of the steam rate, and the enthalpy of the inlet steam (2378 kJ/kg). The total electrical power is the sum of the regeneration and leaching power requirements.

Base Case 3 does not require any seed regeneration plant. Calculations, however, were carried out to determine the minimum electrostatic precipitator efficiency for stack-gas cleaning. These calculations are presented in Subappendix AA 9.1.7.

A 9.1.4 Final Results

The MHD coal requirement, regeneration coal requirement, electric power requirement, makeup seed requirement, cost, and minimum electrostatic precipitator efficiency requirement for all three base cases and their attendent variations are presented in Tables A 9.1.3, A 9.1.4, and A 9.1.5. The percent recovery of seed, the percent ash carry-over, the cost per kilowatt of generating capacity, and the kilojoules per kilowatt are also tabulated.

The per kilowatt costs or energy requirements are generally not affected by the plant size (Points 1, 2, and 3 for Base Cases Number 1 and 2). The costs and energy requirement per kilowatt are greatly affected by the sulfur content of the coal being used (compare Points 5, 6, 7, and 8 for Base Cases 1 and 2 with the other points). The very high-sulfur bituminous coal required a much larger plant and was, therefore, more expensive to build. It required much more reducing gas, thus making the plant less efficient. These costs and energy requirements are not directly comparable with the scrubbers presently in use, since these costs do not include the cost of the electrostatic precipitator for the stack, contingency fees, land costs, and so on; and the energy requirements do not include those of the electrostatic precipitator. At this point, however, it is possible to conclude that, for the bituminous coal, the energy requirements for this seed regeneration system (at least 5.5 to 6.5% of

the total power input) is much higher than for a scrubbing operation (2 to 3% of the total power input)(Reference 9.8). Since the energy requirements of this system are proportional to sulfur flow rate, however, it has an advantage over scrubbers when lower sulfur coals (subbituminous or lignite) are used. Another advantage this system offers is in the form of the waste product: a wet scrubber produces wet calcium sulfate, and the seed regeneration system produces sulfur. The sulfur product is much easier to handle and dispose of than is calcium sulfate.

In Points 11 and 12 of Base Cases 1 and 2, respectively both the cost and energy requirement is much greater than other variations using the same coal. This is due to the addition of a leaching plant to separate the seed and ash mixture and indicates that an initial separation of the ash from the product gases in the combustor is much more desirable than separation in the seed regeneration system.

Seed recovery factors were in the range of 95 to 99%. For a given coal, the seed recovery decreased with an increase in the amount of ash carry-over. The exceptions to this were the cases of 100% ash carry-over, where the ash and seed were separated in the seed regeneration system, and all the seed injected in the second stage of the combustor. Base Case 3 did not require a seed regeneration system and had no ash in the fuel. Seed recovery for this case was 97.47%.

A 9.1.5 Uncertainties

The piece of equipment about which there is the greatest uncertainty is the regenerator, because of the lack of data for the combined reactions A 9.1.1 and A 9.1.2 (which are repeated here for the convenience of the reader):

$$4H_2 + K_2SO_4 \rightarrow 2KOH + H_2S + 2H_2O \qquad \text{(A 9.1.1)}$$

$$CO_2 + 2KOH \rightarrow K_2CO_3 + H_2O \qquad \text{(A 9.1.2)}$$

That the first reaction takes place is known, and there are data to support it. It is not known, however, if the KOH formed on the inside of the particles will be able to contact carbon dioxide and react because

the reaction conditions call for temperatures around 1048°K (1427°F), considerably above the melting point of KOH [633°K (680°F)]. This could lead to massive agglomeration of the small particles (thereby increasing the required time for the diffusion of carbon dioxide into the particle), collection on the walls of the regenerator or heat exchangers, or very slow internal diffusion of carbon dioxide due to the formation of a tight potassium carbonate crust on the particles. If agglomeration and sticking problems occur, the use of an internally cooled reactor in the configuration outlined above would be impractical. Instead, a moving-bed type of reactor might be used. This would require: 1) separation of the ash from the seed prior to its introduction into the ash-potassium sulfate holding tanks; 2) pelletizing the potassium sulfate; and 3) pulverizing the converted seed before injecting it into the combustor.

Another unknown is the degree of agglomeration which will take place in the ash-potassium sulfate holding tanks. Any agglomeration will tend to increase the required residence time in the regenerators and decrease the amount of separation obtainable in ash cyclones between the ash and regenerated seed. If the desired separation is not obtainable in the ash cyclones, then the cyclones will be omitted and a leaching system added to the outlet of the electrostatic precipitator. This modification would increase the power requirement of the system and the cost, but would decrease the amount of makeup potassium carbonate required. This is because the loss of potassium compounds is 10 times greater if reinjected with ash in the first stage of the combustor rather than directly into the second stage. In the present regeneration system, the power requirement is the most critical of the three considerations, and the present cyclone-precipitator separation system is, therefore, more desirable than a precipitator-leaching system. Subappendix AA 9.1.8 presents a comparison between the power usage of a cyclone-precipitator and a precipitator-leaching system.

The gasifier cost estimate was based upon a theoretical design study for an airblown system. Since, at this time, there are no data available to confirm the correctness of this design and the modification to an oxygen-blown system, the gasifier costs are uncertain. Since the

cost of the gasifier is only about 17% of the entire cost, however, a change of 100% in the gasifier would only increase the cost of the overall system by about 17%.

This study has proposed that U-tube heat exchangers and shell and tube steam generators be used to lower the temperature of the gas from 1070 to 541°K (1466 to 514°F). The use of these components depends on the physical characteristics of the regenerated seed and ash in the regenerator gas stream. It was assumed in this study that the solids do not cake on tube surfaces. If this assumption is shown to be incorrect by future data, then three alternatives suggest themselves. First, the gas stream can be cleaned at high temperature. This option would require an increase in the size, and cost, of the solids removal system. Second, a recycle quench with cold gas could be used. The disadvantages of this method are the increase in size of the solids removal and gas cooling system and the lowering of the potential of the energy in the gas. A third option would be to design the exchangers to minimize deposition, if possible. Of the three options, the third is the most desirable and the second the least. It would be expected, however, that the use of any of these options would, at the least, increase the costs of the system.

A 9.1.6 Recommendations for Experimental Work

The following experimental work is necessary either to verify or to provide initial data necessary in the design of several key items in this system.

- Determination of the kinetics of the continued reactions:

$$4H_2 + K_2SO_4 \rightarrow 2KOH + H_2S + 2H_2O \qquad \text{(A 9.1.1)}$$

$$CO_2 + 2KOH \rightarrow K_2CO_3 + H_2O \qquad \text{(A 9.1.2)}$$

Investigate the effects of the percent conversion, particle size, and reaction temperatures on the rate of conversion and the physical characteristics of the potassium sulfate -

potassium hydroxide - potassium carbonate materials. Investigate the effect of these various factors on the agglomeration of the seed material and its collection on reactor surfaces. If these studies determine that the present regenerator design is unsuitable, then a new system design, based on a moving-bed reactor with ash-seed separation performed immediately after removal from the stack, should be considered.

- The amount and rate of agglomeration of ash-seed particles of different initial sizes (0.5 to 20 μm), and different compositions, storage times, pressures, and temperatures in storage bins and gas streams.

- Determination of the amount, rate, and characteristics of the seed materials which cling to the internals of the heat exchange equipment. An evaluation of the effectiveness of methods currently used to remove these deposits is required. If these methods are not satisfactory, then new techniques must be developed, or different systems or heat exchange designs developed to avoid the problems.

A 9.1.7 Conclusions

The use of a potassium sulfate-to-carbonate conversion scheme to remove sulfur from the open-cycle MHD combustion products has been presented. A lack of fundamental experimental data on which to base a design gives rise to serious questions about the operability of the present design. Other alternatives are available, but they too lack a firm experimental basis. Energy requirements for this system are such that the use of any design requiring potassium sulfate-to-carbonate conversion must be seriously questioned. For high-sulfur coal, energy requirements at least 2 to 3 times above that currently achievable with wet scrubbers seems unavoidable.

The ease of handling the waste product of this system (sulfur), as compared to that of a wet scrubber system (calcium sulfate or magnesium sulfate added) provides incentive to continue the development of this system.

Subappendix AA 9.1.1

COST AND ENERGY COMPARISON OF ELECTROLYSIS
AND COAL GASIFICATION FOR HYDROGEN PRODUCTION

To reduce one mole of potassium sulfate according to the reaction in Equation AA 9.1.1.1

$$4H_2 + K_2SO_4 \rightarrow 2KOH + 2H_2O + H_2S \qquad \text{(AA 9.1.1.1)}$$

four moles of hydrogen are necessary. Using presently available electrolysis methods (Reference 9.9), the lowest energy requirement per kilogram mole of hydrogen is about 105.6 kWh/kg mole hydrogen, or 75% efficiency. If the power to produce the electricity to drive the electrolysis units can be obtained from a 50% efficient MHD plant, then the total efficiency drops to around 37.25%. In comparison, gasifier efficiencies range from 85% (Reference 9.10) to 93% (Reference 9.11). The use of a gasifier to produce the needed hydrogen assumes that the unused gases and enthalpy of the gases can be used elsewhere in the plant.

Present estimates of the costs of electrolysis units are about \$35.64 x 10^6/kg hydrogen/s (\$4,500 lb/hr). For an MHD plant using 160 kg/s of 3.9% sulfur coal, and removing 83.3% of the sulfur by use of potassium carbonate, an electrolysis unit to produce the necessary hydrogen would cost at least \$49.2 million. A comparable coal gasifier (oxygen-blown) would cost \$20.6 million.

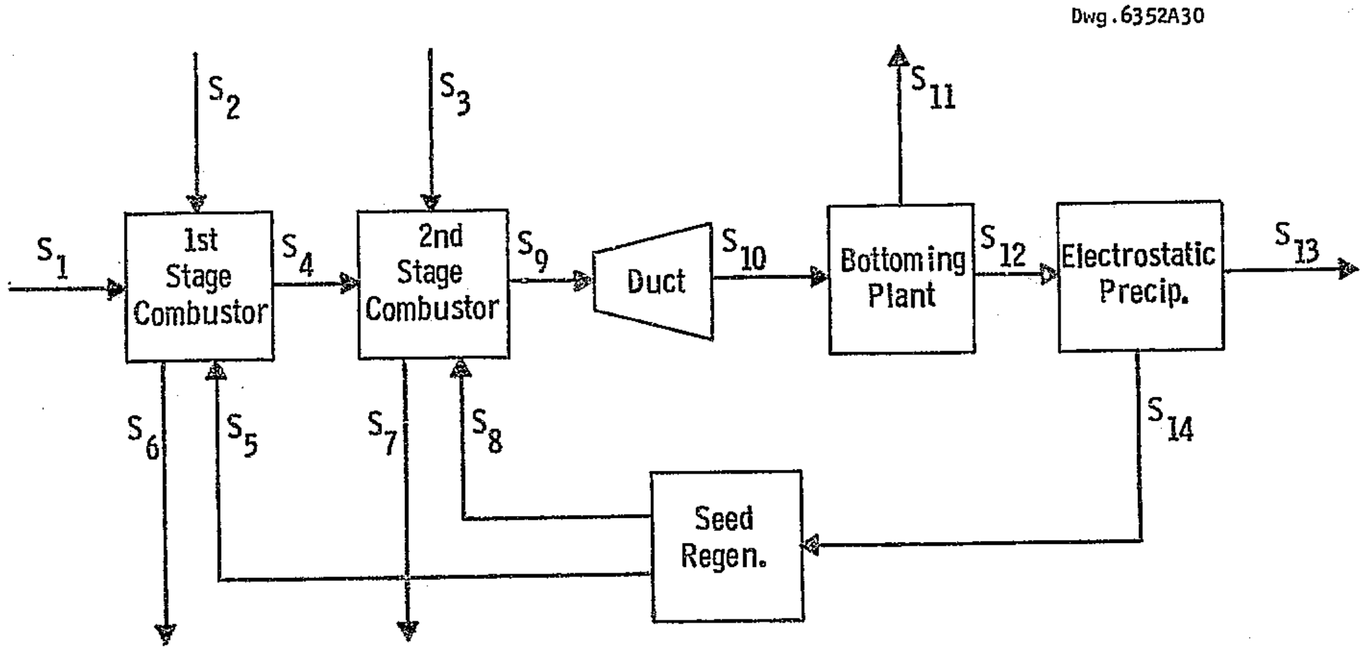
Subappendix AA 9.1.2

OVERALL POTASSIUM BALANCE ON MHD SYSTEM FOR BASE CASE 2

The following material balance is for Base Case 2, where the coal is burned in the combustors directly. For this case, the following assumptions were made concerning potassium loss from the system:

1. 20% (Reference 9.7) of the potassium entering the first stage of the combustor would be rejected in the slag as potassium carbonate. Of this rejected potassium, 50% of the potassium carbonate will react with sulfur to form potassium sulfate.
2. 2% (Reference 9.7) of the potassium introduced into the second stage of the combustor is lost to the slag.
3. 0.01% of the potassium is lost in the bottoming plant.

A flow sheet of the overall system is presented in Figure AA 9.1.2.1. The S values correspond to the total mass flow rate; the X_y values to the mass fraction of the compound of element "y". A number subscript refers to a particular flow to or from a vessel was indicated in Figure AA 9.1.2.1. For a coal-ash feed rate of S_1, $X_{K,1}$ mass fraction will be potassium. From the British work (Reference 9.4), the ash which comes out of the electrostatic precipitator and is returned to the first-stage combustor at a flow rate, S_5, has about 60%, by weight, potassium sulfate. Therefore, for an ash mass fraction of $X_{ash,5}$ the amount of potassium returned with the ash to the first stage of the combustor is $0.6\ X_{ash,5}\ S_5$. Carrying out a mass balance on the first-stage combustor for potassium, one obtains Equation AA 9.1.2.1:



S_1	Coal + Ash
S_2	Air
S_3	K_2CO_3 Makeup
S_4	1st - 2nd Stage Gas
S_5	Ash, K_2SO_4, K_2CO_3
S_6	Ash, K_2SO_4, K_2CO_3
S_7	K_2SO_4, K_2CO_3
S_8	K_2SO_4, K_2CO_3
S_9	Gas + K_2SO_4 + Ash
S_{10}	Gas + K_2SO_4 + Ash
S_{11}	Ash + K_2SO_4 + Ash
S_{12}	Gas + K_2SO_4 + Ash
S_{13}	Gas + K_2SO_4 + Ash
S_{14}	K_2SO_4 + Ash

Fig. AA 9.1.2.1—Flow sheet of open-cycle MHD system

$$X_{K,4}\,S_4,\ \text{kg/s} = X_{K,1}\,S_1 + 0.6\,X_{ash,5}\,S_5 - 0.2(X_{K,1}\,S_1 + 0.6\,X_{ash,5}\,S_5) =$$

$$0.8\,X_{K,1}\,S_1 + 0.48\,X_{ash,5}\,S_5 \qquad \text{(AA 9.1.2.1)}$$

One notes that the potassium rejected with the ash is assumed to use only 50% of its available potassium carbonate for sulfur removal, the reasoning for which follows. The British work (Reference 9.4) has shown that when the slag and gas are separated at temperatures above about 2200°K (1478°F), one should expect little or no potassium in the slag. If one injects the ash-potassium mixture, it is more reasonable to assume that the 20% potassium carried over in the slag never left the ash-potassium mixture, and that only the potassium carbonate near or on the surface reacted with sulfur. Since no data were available, therefore, it was assumed that 50% of the potassium available as potassium carbonate and carried over in the slag reacted to form potassium sulfate.

If Y_S is the fraction of the sulfur which must be removed from the coal to meet current federal standards, then the sulfur removed is given by the mass balance in Equation AA 9.1.2.2:

$$(X_{S,1})(S_1)(Y_S) - [(0.5)(0.2)(0.6)(X_{ash,5})(S_5) + (0.5)(0.2)(X_{K,1}\,S_1)]\,\frac{32}{78}$$

$$= (X_{S,1})(S_1)(Y_S) - [(0.06)(X_{ash,5}\,S_5) + (0.1)(X_{K,1}\,S_1)]\,0.140$$

$$\text{(AA 9.1.2.2)}$$

The next step is to calculate the amount of sulfur to be removed in the second stage of the combustor. Assuming that Y_c is the fraction of all the potassium sulfate which enters at a flow rate, S_{14}, that is converted to potassium carbonate, then the potassium-sulfur balance is as given by Equation AA 9.1.2.3:

$$\frac{0.98}{78}(S_3 + Y_c\, S_8\, X_{K,8}) + \frac{0.8\, X_{K,1}\, S_1}{78} + \frac{0.48\, X_{ash,5}\, S_5\, Y_c}{78}$$

$$= \frac{X_{S,1}\, S_1\, Y_S - (0.06\, X_{ash,5}\, S_5 + 0.1\, X_{K,1}\, S_1)\, 0.410}{32}$$

or

$$S_3,\ \text{kg/s} = S_1\,(2.49\, X_{S,1}\, S_1\, Y_S - 0.714\, X_{K,1}) - Y_c\, S_8\, X_{K,8}$$

$$- 0.55\, X_{ash,5}\, S_5\, Y_c \qquad \text{(AA 9.1.2.3)}$$

where S_3 is the potassium flow rate in the form of potassium carbonate.

The next requirement to be met in the second stage of the combustor is that the potassium be 1% by weight of the total mass in S_8, excluding the ash. This requirement is reflected in Equation AA 9.1.2.4:

$$\frac{0.98(S_3 + X_{K,8}\, S_8) + 0.8\, X_{K,1}\, S_1 + 0.48\, X_{ash,5}\, S_5}{S_1(1 - X_{ash,1}) + S_2 + 0.98(S_3 + Y_c\, X_{K,8}\, S_8)\,\frac{138}{78} + 0.48\, X_{ash,5}\, S_5\, Y_c\,\frac{138}{78}} =$$

$$0.01 + 0.98(1 - Y_c)\, X_{K,8}\, S_8\, \frac{174}{78} + 0.48\, X_{ash,5}\, S_5(1 - Y_c)\,\frac{174}{78}$$

(AA 9.1.2.4)

Solving for $S_8\, X_{K,8}$, one obtains

$$X_{K,8}\, S_8 =$$

$$\frac{-S_5\, X_{ash,5}(0.0022\, Y_c + 0.469) + 0.01\, S_2 + 0.963\, S_3 + S_1(0.01(1 - X_{ash,1}) - 0.8\, X_{K,1})}{(0.0046\, Y_c + 0.958)}$$

(AA 9.1.2.5)

If Y_p is the mass fraction of the ash which must be captured before the flue gas meets current federal standards on particulate emission and is allowed up the stack, then the amount of particulate matter to be allowed up the stack is $(1 - Y_p) X_{ash,1} S_1$ kg/s. This mass rate of particulates would, in the present case, include any potassium sulfate allowed to escape to the environment. If Y_E is the fraction of solids in S_{12} allowed to leave in S_{13}, then

$$Y_E S_{12} X_{solids,12} = (1 - Y_p) X_{ash,1} S_1$$

or

$$Y_E \left[(0.98)\left(\frac{174}{78}\right)(S_3 + S_8 X_{K,8}) + \frac{174}{78}(0.8 X_{K,1} S_1 + 0.48 X_{ash,5} S_5) + X_{ash,1} S_1 + X_{ash,5} S_5) Y_I\right] = (1 - Y_p) X_{ash,1} S_1$$

where Y_I is the mass fraction of the ash carried over from the combustor. Solving for Y_E one obtains

$$Y_E = \frac{(1 - Y_p) X_{ash,1} S_1}{2.9(S_3 + S_8 X_{K,8}) + S_1(1.78 X_{K,1} + Y_I X_{ash,1}) + X_{ash,5} S_5(1.07 + Y_I)} . \qquad \text{(AA 9.1.2.6)}$$

If a mass balance is done based on the ash entering and leaving the regeneration system, one obtains

$$X_{ash,5} S_5 = (1 - Y_E)(X_{ash,1} S_1 + X_{ash,5} S_5) Y_I$$

or, solving for $X_{ash,5} S_5$, one obtains

$$X_{ash,5} S_5 = \frac{(1 - Y_E) X_{ash,1} S_1 Y_I}{1 - (1 - Y_E) Y_I} \qquad \text{(AA 9.1.2.7)}$$

Substituting Equation AA 9.1.2.7 into Equations AA 9.1.2.3, AA 9.1.2.5, and AA 9.1.2.6, one obtains

$$S_3 = S_1 \left(2.49\, X_{S,1}\, Y_5 - 0.714\, X_{K,1}\right) - Y_c\, S_8\, X_{K,8}$$

$$- \frac{(1 - Y_E)\, X_{ash,1}\, S_1\, Y_I}{(1 - (1 - Y_E)\, Y_I)} (0.061 + 0.49\, Y_c) \qquad \text{(AA 9.1.2.8)}$$

$$X_{K,8}\, S_8 = \frac{\left[\begin{array}{c} S_1(0.01\,(1 - X_{ash,1}) - 0.8\, X_{K,1}) + 0.01\, S_2 - 0.963\, S_3 \\ - \dfrac{(1 - Y_E)\, X_{ash,1}\, S_1\, Y_I\, (0.0022\, Y_c + 0.469)}{(1 - (1 - Y_E)\, Y_I)} \end{array}\right]}{(0.0049\, Y_c + 0.958)}$$

(AA 9.1.2.9)

$$Y_E = \frac{(1 - Y_p)\, X_{ash,1}\, S_1}{\left[\begin{array}{c} 2.9(S_3 + S_8\, X_{K,8}) + S_1(1.78\, X_{K,1} + Y_I\, X_{ash,1}) \\ + \dfrac{(1 - Y_E)\, X_{ash,1}\, S_1\, Y_I (1.07 + Y_I)}{(1 - (1 - Y_E)\, Y_I)} \end{array}\right]}$$

(AA 9.1.2.10)

Substituting for S_3 in Equations AA 9.1.2.9 and AA 9.1.2.10 using Equation AA 9.1.2.8, one obtains

$$X_{K,8}\, S_8 = \frac{(1 - Y_E)\, X_{ash,1}\, S_1\, Y_I}{(1 - (1 - Y_E)\, Y_I)} \left(\frac{0.47\, Y_c - 0.41}{0.958(1 - Y_c)}\right)$$

$$+ \frac{0.01\, S_2 - 2.4\, X_{S,1}\, S_1\, Y_S + 0.034\, S_1\, X_{K,1} + 0.01(1 - X_{ash,1})\, S_1}{0.958(1 - Y_c)}$$

(AA 9.1.2.11)

and

$$Y_E = \frac{(1 - Y_p)\, X_{ash,1}\, S_1 (1 - (1 - Y_E)\, Y_I)}{(1 - (1 - Y_E)\, Y_I)(S_8\, X_{K,8}\ 2.19 (1 - Y_c) + 5.45\, X_{S,1}\, S_1\, Y_S - 0.223\, X_{K,1}\, S_1)}$$

$$+ (0.963 - 1.07\, Y_c (1 - Y_E) + 1)\, X_{ash,1}\, S_1\, Y_I \qquad \text{(AA 9.1.2.12)}$$

Substituting for $X_{K,8}\, S_8$ from Equation AA 9.1.2.11 into Equation AA 9.1.2.12, one obtains

$$Y_E = \frac{(1 - Y_p)}{\dfrac{Y_I (1 + 0.026 (1 - Y_E))}{(1 - (1 - Y_E)\, Y_I)} + \dfrac{0.0229 (S_2 + S_1)}{X_{ash,1}\, S_1} - 0.0229 - 0.145\, \dfrac{X_{K,1}}{X_{ash,1}}}$$

(AA 9.1.2.13)

If the only losses of potassium from the system are through the combustor stages, the steam bottoming plant, and the stack, than the percent recovery is

% Recovery =

$$100 \left[1 - \frac{S_3}{0.98\ (S_3 + X_{K,8}\, S_8) + X_{K,1}\, S_1 + 0.6\ (78/94)\, X_{ash,5}\, S_5} \right].$$

(AA 9.1.2.14)

As these equations are solved for various Y_c's, a point will be reached where S_3, the potassium flow rate into the system, will equal the flow of potassium out of the system. This balance is written as

$$S_3 = 0.02\ (S_3 + X_{K,8}\ S_8) + (0.2)(0.6)\ X_{ash,5}\ S_5$$

$$+\ 0.0001\ (0.98\ (S_3 + X_{K,8}\ S_8) + (0.8)(0.6)\ X_{ash,5}\ S_5)$$

$$+\ Y_E\ (0.9999\ (0.98\ (S_3 + X_{K,8}\ S_8) + (0.8)(0.6)\ X_{ash,5}\ S_5)) - 0.8\ X_{K,1}\ S_1$$

or

$$S_3 = (0.02 + 0.98\ Y_E)(S_3 + X_{K,8}\ S_8)$$

$$+\ X_{ash,5}\ S_5\ (0.12 + 0.48\ Y_E) - 0.8\ X_{K,1}\ S_1 \tag{AA 9.1.2.15}$$

Subappendix AA 9.1.3

EQUIPMENT SIZING AND COSTS

AA 9.1.3.1 Potassium Sulfate-Ash Storage-Feed System

Potassium Sulfate Surge Bin

Flow rate of potassium sulfate in = 32.71 kg/s

Flow rate of ash in = 3.84 kg/s

Total flow in = 36.55 kg/s

Assuming holdup time of 3.6 ks (1 hr) and density of potassium sulfate and ash powder to be 190 kg/m^3 (Reference 9.4),:

$$\text{Storage volume required, } m^3 = \frac{36.55 \text{ m} \cdot 3600 \text{ s}}{190} = 691$$

Temperature of storage ≈ 350°K

Pressure of storage ≈ 101.3 kPa

The vessel chosen was 4.57 m (15 ft) in diameter by 14.02 m (46 ft) high and weighed 100 Mg (10.23 tons). Installed cost was estimated to be $120,000 for each of three vessels. [Note: Pressure vessel and bin weights and cost were estimated from Guthrie (Reference 9.12).] Vessel weights include supports, heads, shells, and contents. This vessel requires a shaking device to aid in transferring the contents. Carbon steel is specified for all material. The price given is adjusted to early 1974 by using CE Plant Cost Index (Reference 9.13).

Potassium Sulfate Locks. Each lock fills in 89 s. Pump time allowed is 178 s. The temperature of operation is about 350°K (170°F). The pressure of operation is 0.1013 to 1.722 MPa (1 to 17 atm). Locks are pressurized utilizing hot fuel gas. Vessel size is chosen to be 2.44 m (8 ft) diameter by 3.66 m (12 ft) high. A shaking device to aid in transferring the contents is required. The vessel weight was calculated to be 13.337 Mg (14.701 tons). The installed vessel cost was estimated to be $80,000 based on construction with carbon steel.

AA 9.1.3.2 Coal Gasifier System

This system is a scaled version presented by Vidt and Peterson, Reference 9.11, utilizing an oxygen-blown gasifier rather than an air-blown gasifier. The scale factor was obtained as follows. The number of kilogram moles of potassium sulfate to be converted is:

$$\text{kg moles } K_2SO_4 = (32.71 \frac{\text{kg}}{\text{s}} \; K_2SO_4) \left(\frac{1 \text{ kg mole } K_2SO_4}{174 \text{ kg } K_2SO_4}\right) (0.869)$$

(0.869 is the required conversion of potassium sulfate to potassium carbonate). Assuming that at least 50% of the available carbon monoxide and hydrogen is used (Reference 9.6) in converting the potassium sulfate to the potassium carbonate, and that these are the limiting reagents,

$$\text{kg moles of gas} = \frac{(0.169 \frac{\text{kg mole}}{\text{s}} \; K_2SO_4)\left(\frac{4 \text{ kg mole } (H_2 + CO)}{\text{kg mole } K_2SO_4}\right)}{(0.50)(0.6625) \frac{\text{kg mole } (H_2 + CO)}{\text{kg mole gas}}}$$

$$\text{kg moles of gas} = 2.04$$

The molecular composition of the gas is based upon data presented by Hamm (Reference 9.14) for Illinois No. 6 bituminous, and modified for an oxygen-blown system. For a gas of molecular weight 21.3, the gas rate is

$$2.04 \left(\frac{\text{kg mole gas}}{\text{s}}\right)(21.3) = 43.45 \frac{\text{kg}}{\text{s}}$$

The coal rate is

$$\frac{43.45 \frac{\text{kg}}{\text{s}} \text{ gas}}{2.22 \frac{\text{kg gas}}{\text{kg coal}}} = 19.55 \frac{\text{kg coal}}{\text{s}} \text{ (dried)}$$

The oxygen rate is

$$\left(19.55 \frac{\text{kg}}{\text{s}} \text{ coal}\right)\left(0.726 \frac{\text{kg } O_2}{\text{kg coal}}\right) = 14.20 \frac{\text{kg } O_2}{\text{s}}$$

The steam rate is

$$\left(19.55 \frac{\text{kg}}{\text{s}} \text{ coal}\right)\left(0.608 \frac{\text{kg steam}}{\text{kg coal}}\right) = 11.89 \frac{\text{kg}}{\text{s}} \text{ steam}$$

Since Vidt and Peterson's fuel gas rate was 140 kg/s, the required plant size is scaled as follows:

$$\frac{43.45}{140} = 31\% \text{ of size for gas-handling equipment,}$$

and

$$\left(\frac{4.28}{2.22}\right)\left(\frac{43.45 \frac{\text{kg gas}}{\text{s}}}{140 \frac{\text{kg gas}}{\text{s}}}\right) = 0.598 \text{ the size for solids equipment.}$$

Coal Bin. The coal bin was scaled down from Vidt and Peterson to a volume of 8.15 m^3 (287.8 ft^3). Carbon steel construction was assumed. Dimensions were 3.66 by 3.66 m (11.81 by 11.81 ft), with the two sides

sloped at 60° to horizontal to form the chute. It was found from costing other nonpressure items that the current installed cost was twice the cost reported by Vidt and Peterson. A cost of $50,000 was assumed. The weight was about 154.682 Mg (170.50 tons) (full).

Coal Reclaimer Conveyor. A capacity of 29.356 kg/s (116.24 ton/hr) was required. It was a belt-type conveyor 0.362 m (14.25 in) wide, 24.1 m (79 ft) long, and 10.4 m (34.1 ft) high. From pricing other mechanical equipment (pumps), Vidt and Peterson's price was found to be multiplied by 4 (due to price increases). The installed price of $100,000 was assumed. Assuming a bulk density of around 803 kg/m^3 (50 lb/ft^3), and obtaining a value of 0.865 kW/3.05 m (1.16 hp/10 ft) lift, and 0.865 kW/30.5 m (1.16 hp/100 ft) centers from Perry (Reference 9.15), a power requirement of 3.81 kW (5.1 hp) is obtained. Allowing a margin of 100%, this works out to about 7.46 kW (10 hp).

Coal Crusher Surge Bin. A scaled size was found to be 3.66 m (12 ft) diameter by 10.67 m (35 ft) high. The full weight of this bin, which was made from carbon steel, was 116 Mg (127.87 tons). The cost, estimated from Guthrie, was $90,000.

Coal Crusher/Dryer. The Illinois No. 6 bituminous coal was assumed to have been dried from 13% moisture to 3% moisture before being used in the gasifier. At a coal rate of 23.61 kg/s and an efficiency of 50%, the amount of energy required is

$$\left(23.61\ \frac{kg}{s}\ \text{coal}\right)\left(\frac{10\%\ \text{removal}}{50\%\ \text{efficiency}}\right)\ 540\ \frac{kcal}{kg} = 10.66\ \text{MWt}$$

A cost estimate based on the same assumption as the coal reclaimer conveyor is $1,280,000. The weight is 116 Mg (127.87 ton) and the power requirement is about 298 kW (400 hp) (Reference 9.15).

Coal Surge Bins. Two coal surge bins are required. Each is 3.66 m (12 ft) diameter and 18.59 m (61 ft) high. They were constructed of carbon steel. Loaded weight of each is 145 Mg (159.83 tons), and the cost is $210,000.

Predried Coal Elevator. This is a bucket elevator with a 15.24 m (50 ft) lift. The capacity is 29.293 kg/s (116.24 ton/hr). Power requirements are estimated at about 11.19 kW (15 hp) Reference 9.15 (25% over estimate). The cost is calculated at $70,000.

Sized Coal Feedlocks. These vessels are 1.83 m (6 ft) diameter and 3.66 m (12 ft) high. They are constructed of carbon steel and built to withstand pressures up to 2.026 MPa (20 atm). The design is based on 0.833 cycles/ks (3 cycles/hr). The weight when filled was 18.727 Mg (20.642 ton). Cost was estimated at $60,000 each. Two were required.

Sized Coal Feed Hoppers. Same specifications as on the sized coal feedlocks.

Coal Preheater. This vessel is 2.74 m (9 ft) diameter and 4.27 m (14 ft) high. It was constructed of carbon steel to withstand pressures up to 2.026 MPa (20 atm) at a temperature of 617°K (651°F). Internals include 2 two-stage Ducon cyclones, 0.61 m (2 ft) diameter and 1.2 m (3.94 ft) high. The weight was estimated at 14.091 Mg (15.532 ton) and the cost at $80,000.

Volatilizers. These vessels are lined with 10.16 cm (4 in) of Harbison Walker Castolast G and 1.22 m (4 ft) of Harbison Walker Castable. The upper part of the stainless steel clad shell is 5.45 m (17.88 ft) dimater and 3.73 m (12.23 ft) high. The lower section is 2.48 m (8.13 ft) diameter and 9.44 m (30.97 ft) high. The design pressure of the shell is 2.026 MPa (20 atm). The design temperature is 617°K (651°F) and that of the internals is 1144°K (1600°F). The weight is 166.344 Mg (183.36 ton) and the estimated cost is $680,000 each. Internals include a refractory partition and 2 Incoloy 800 Ducon cyclones with dip legs and flapper valves. Two units are required.

Gasifiers. These vessels are constructed of the same materials as the volatilizers. The upper section is 5.46 m (17.91 ft) diameter and 2.48 m (8.14 ft) high and the lower section is 2.48 m (8.14 ft) diameter and 7.19 m (31.72 ft) high. Internal design temperature is 1367°K (2000°F). The weight is estimated at 132.683 Mg (146.25 ton) and cost at $520,000 each. Two units are required.

Ash Quench Pots. There are two vessels, each stainless steel clad. They are 1.83 m (6 ft) diameter and 3.66 m (12 ft) high. Pressures up to 2.026 MPa (20 atm) and temperatures to 617°K (651°F) can be handled. Full weights are 8.273 Mg (9.119 ton) and the cost is estimated at $60,000 each. Two units are required.

Ash Slurry Locks. Same specification as ash quench pots.

Ash Slurry Pumps. These are two 8.2 kg/s, 7.48 kW rubber-lined mud pumps. The medium they pump is 18% ash and water. The cost is estimated at $10,000 each. Two are required.

AA 9.1.3.3 Regeneration and Separation System

The regeneration process is based on data presented by the United States Bureau of Mines (USBM). In their work, cylinders of potassium sulfate plus ash 0.310 cm (0.122 in) diameter and 0.64 cm (0.252 in) high were used. In this MHD study, the size of the potassium sulfate particles was taken to be 2 μm.

From Levenspiel (Reference 9.16) for a first-order reaction with the rate of gas diffusion through a particle being the rate controlling factor, Equation AA 9.1.3.1 results:

$$T = \frac{C_{K_2CO_3} R^2}{1.5 D_{H_2} C_{H_2}} \left[1 - 3\left(\frac{r_c}{R}\right)^2 + 2\left(\frac{r_c}{R}\right)^3\right] \qquad \text{(AA 9.1.3.1)}$$

where R = the outside radius of the particle

r_c = the final radius of the unreacted core

$C_{K_2CO_3}$ = the density of the potassium carbonate shell

D_{H_2} = the diffusion coefficient of hydrogen in potassium carbonate

C_{H_2} = is the concentration of hydrogen in the gas.

Aris (Reference 9.17, Figure 6.7) shows that there is little difference between the reaction rate and cylinder of the same diameter. Therefore, the USBM data will be assumed also to apply directly to a spherical geometry. For a final conversion of X, (X = 0 is no conversion, X = 1 complete conversion), the relation between the unreacted core radius (r_c) a d the final conversion is

$$X = 1 - \frac{r_c^3}{R^3}$$

or

$$\frac{r_c}{R} = (1 - X)^{1/3} \qquad \text{(AA 9.1.3.2)}$$

Substituting Equation AA 9.1.3.2 into Equation AA 9.1.3.1 for r_c/R, one obtains

$$T = \frac{C_{K_2CO_3} R^2}{1.5\ D_{H_2}\ C_{H_2}} \left[1 - 3\left(1 - X\right)^{2/3} + 2\left(1 - X\right)\right] \qquad \text{(AA 9.1.3.3)}$$

The time rates for two different values of C_{H_2} and X is given in Equation AA 9.1.3.4

$$\frac{T_1}{T_2} = \frac{C_{H_{2(2)}}\ R_1^2}{C_{H_{2(1)}}\ R_2^2} \left[\frac{1 - 3(1 - X_1)^{2/3} + 2(1 - X_1)}{1 - 3(1 - X_2)^{2/3} + 2(1 - X_2)}\right] \qquad \text{(AA 9.1.3.4)}$$

In Reference 9.6, $C_{H_2} = 1$, $X = 0.5$, $R = 0.16$ cm (0.062 in), and $T_1 = 4.320$ ks (72 min). In the present study, $C_{H_2} = 0.21$, $X = 0.9$, and $R = 1$ µm. Solving Equation AA 9.1.3.4 for these values, one obtains

$$T_1 = (72)(60)\left(\frac{1}{0.21}\right)\left(\frac{1 \times 10^{-4}}{0.16}\right)^2\left(\frac{1 - 3(1 - 0.9)^{2/3} + 2(1 - 0.9)}{1 - 3(1 - 0.5)^{2/3} + 2(1 - 0.5)}\right)$$

$$= 0.0405 \text{ s}$$

for the reactor residence time.

If the particles should agglomerate to about 20 µm diameter, then the reactor residence time would be increased by a factor of 100 to 4.05 s. Assuming the larger particles, and a safety factor of 300%, the residence time is 14.2 s. The gas volume flow rate, $\dot{V}$, was calculated from the perfect gas Equation AA 9.1.3.5 in SI units

$$\dot{V} = \frac{\dot{M}RT}{p} = 20.4 \quad \frac{8.31 \text{ kPa}}{1494 \text{ kPa}} \, 1048 = 11.89 \, \frac{m^3}{s} \qquad \text{(AA 9.1.3.5)}$$

The total volume for regeneration is the product of $\dot{V}$ and the residence time, or 169 m^3.

Regenerators. If two regenerators of 84 m^3 (2966 ft^3) each are used, and a gas velocity of about 1.52 m/s (4.986 ft/s) is maintained then the size will be 3.16 m (10.37 ft) id and 12.5 m (41 ft) high.

The required heat transfer area is 340 m^2 (3660 ft^2). For each 5.08 cm (2 in) diameter tube the area per unit length is 0.1596 m^2/m (1.718 ft^2/ft). If the cross section has a diameter of 3.16 m (10.37 ft) and the tubes are arranged on 7.62 cm (3 in) square pitch, then a maximum of 1350 pipes will fit. For the 1350 pipes the heat transfer area per meter of regenerated length is 216 m^2/m (708.6 ft^2/ft), so two 1 m sections will be required.

The velocity in the unimpeded area [7.84 m^2 (84.38 ft^2)] of the reactor is 1.52 m/s (4.987 ft/s), since the total pipe cross-sectional area is 2.74 m^2 (29.493 ft^2). Neglecting the wall thickness of the tubes, the flow velocity in the tubes would be approximately 4.35 m/s (14.27 ft/s). To increa... the velocity of the process gas-solids mixture through the pipes, the number of pipes in each section will be halved to 675. There will be four 1 m sections of heat transfer area. The velocity of the seed-process gas mixture will be 8.7 m/s (28.54 ft/s) through the pipes. The cost is calculated as follows:

Shell [3.16 m (10.37 ft) dia by 12.5 m (41 ft) high, rated at 2.026 MPa (20 atm), 1200°K (1700°F), 316 SS clad]	$ 670,000
4 Heat Exchangers [133 m^2 (1432 ft^2) area each, rated at 2.026 MPa (20 atm), 1200°K (1700°F), 316 SS]	670,000
3 Interconnecting pipes [1 m dia, 2.026 MPa (20 atm), 6 m (19.68 ft) long]	10,000
8 Flanges [4.136 MPa (600 psi) rating SS]	100,000
6 Elbows (SS, 90°)	40,000
TOTAL	$1,490,000

Ash Removal Cyclone. The purpose of these cyclones was to separate the ash from the gas-seed stream. In all cases except the 100% ash carry-over system, this ash is cycled directly to the first stage of the combustor, as about 30% by weight of the ash is potassium oxide (Reference 9.4). The potassium sulfate contained in this ash is assumed to have been converted to potassium carbonate (86.9% conversion potassium sulfate to potassium carbonate), and to be available to remove sulfur from the gas stream. In the case of 100% ash carry-over, a small leaching, drying plant will be added. This plant is considered later.

The gas stream entering the cyclone will have been cooled to 541°K (514°F) by the preceding heat exchangers. The volume flow rate is

then calculated from the perfect gas law to be 5.75 m^3/s (203.06 ft^3/s) for a gas temperature and pressure of 541°K (514°F) and 1.4702 MPa (14.51 atm), respectively. For this volume flow rate and an assumed pressure drop of 50.65 kPa (0.5 atm) through the regenerators and heat exchangers. Using the equation

$$D_{p_c} = \left[\frac{9\mu\ B_c}{2\pi\ N_c V_c\ (\rho_s - \rho)}\right]^{0.5}$$

[obtained from Perry (Reference 9.15)] where

- D_{p_c} = particle size of which half is removed (taken to be 10×10^{-6} m)
- μ = viscosity of gas stream ($\sim 2.75 \times 10^{-2}\ \frac{g}{m/s}$)
- B_c = duct width in meters (square duct inlet assumed)
- N_c = number of rotations of gas in cyclone (taken as 5)
- V_c = velocity of gas stream into inlet duct (assumed to be 5.75 $m^3/s/B_c^{\ 2}$)
- $\rho_s - \rho$ = density difference between particles and gas (2.20×10^6 g/m^3)

Solving for B_c one obtains

$$B_c = \left(\frac{D_{p_c}^{\ 2}\ 2\pi\ N_c\ 5.75\ (\rho_s - \rho)}{9\mu}\right)^{1/3} = 0.548\ \text{m}$$

Using the standard cyclone proportions presented in Perry (Reference 9.16), the diameter and heights of the straight and conical sections were each 2.19 m (7.185 ft). This vessel is stainless steel clad and rated at

2.026 MPa (20 atm) and 600°K (620°F). The approximate weight is 8.886 Mg (9.795 ton) and the cost is $240,000 each.

The pressure drop can be estimated from Perry (Reference 9.15) from the equation

$$DP_i,\ kPa = 3.99 \times 10^{-3}\ \rho_{gas}\ V_{gas}^{\ 2}$$

where ρ_{gas} in kg/m^3 units and V_{gas} is in m/s units. For ρ = 8.6 kg/m^3 and V_{gas} = 19.15 m/s, the pressure drop is 11.14 kPa (0.11 atm).

Potassium Carbonate Removal. The use of a cyclone to remove the remaining potassium carbonate dust is impractical. The smallest particle which can be removed, before sonic velocity is reached, with 50% efficiency is about 1.00 μm. This means that at best, 95% of the dust will be captured. To overcome this problem, an electrostatic precipitator is proposed. From Heywood (Reference 9.4), the relation between efficiency and size is:

$$E = 1 - e^{(-0.0427\qquad)(A/V)}$$

where A is the area and V is the volume flow per second. For a volume flow rate, V, of 5.79 m^3/s (12,268 ft^3/min), an area, A, of 718.4 m^2 (7733 ft^2) is required for a collection efficiency of 0.9954. If the unit is enclosed in a cylinder 6.1 m (20 ft) diameter, and the plate spacing is 0.203 m (8 in), then 22 plates, each 4.24 m by 7.70 m (13.91 by 25.26 ft) are required. Energy usage, estimated from Perry (Reference 9.15) is 9.2 kW. The cost is estimated to be $1,040,000, installed.

Ash Lockhoppers. The ash rate is 3.84 kg/s (8.466 lb/s). For 1 hr holdup and assuming a density of 190 kg/m^3 (11.86 lb/ft^3), a volume of 72.8 m^3 (2571 ft^3) is needed. These vessels are rated at 2.026 MPa (20 atm) and 600°K (620°F) and are constructed of carbon steel. Their dimensions are 2.74 m (9 ft) diameter by 6.17 m (20.24 ft) high. Two are required. Each weighs about 22.179 Mg (24,447 ton) and costs $160,000.

Potassium Carbonate Lockhoppers. The potassium carbonate and potassium sulfate flow rates are 26.64 kg/s (58.73 lb/s) for 3.6 ks (1 hr) holdup; and, assuming a density of 190 kg/m^3 (11.86 lb/ft^3), a volume of 505 m^3 (17,834 ft^3) is needed. These vessels are rated at 2.026 MPa (20 atm) and 600°K (620°F), and are constructed of carbon steel. Their dimensions are 4.88 m (16 ft) diameter by 13.1 m (42.98 ft) high. Two are required. Each weighs about 98.335 Mg (108.4 ton) and costs $120,000.

AA 9.1.3.4 Sulfur Removal System

The sulfur removal system consists of a Claus Plant, gas heat exchangers to reheat the gas to make it suitable for injection into the combustor, and turbine and compressor units to lower the fuel gas to 0.606 MPa (6 atm), and produces 1.722 MPa (17 atm) oxygen and fuel gas for the gasifiers and lockhoppers, respectively. The heat exchangers are also used to cool the gas feed into the cyclones.

Claus Reactor. Previous work in this field (Reference 9) has established that a reactor residence time of 1 s is sufficient. It is assumed that the reactor has a total of half its volume filled by solid catalyst. The catalyst is assumed to weight 1.602 Mg/m^3 (100 lb/ft^3). The volume flow rate of gas, Vol, to be treated is

$$\mathrm{Vol} = \left[(5.79\ m^3/s + \frac{(0.099)(1.88)(0.082)(541)}{(2)(14.39)} \right] = 6.028\ \frac{m^3}{s}$$

The 0.099 is the mole fraction of hydrogen sulfide in the gas stream, Half of this amount of sulfur dioxide is needed, at a temperature of 541°K (514°F), 1.458 MPa (14.39 atm), and 1.88 moles of gas inlet. The total volume of the reactor is therefore 12.16 m^3 (429.4 ft^3). This vessel is a stainless steel clad pressure vessel rated at 2.026 MPa (20 atm) and 600°K (620 °F). The size is 1.6 m (5.25 ft) diameter and 6 m (19.69 ft) high. The total weight is 14.075 kg (0.5.517 ton) and the cost is $210,000 [assuming a catalyst cost of $2.20/kg ($1.00/lb)].

Scrubber/Demister. The lower portion of this vessel is an open spray tower in which the gaseous sulfur is condensed, solidified, and washed from the gas stream. The upper part of the tower contains wire mesh demisters. The maximum gas velocity is 1 m/s (3.28 ft/s) in the scrubber. This gives a vessel diameter of 2.79 m (9.15 ft). For a contact time of 5 s, the scrubber height is 5 m (16.40 ft). The water temperature used in the scrubber is 332°K (138°F). The gas is assumed to reach this temperature. The volume flow rate in the demister section for a flow rate of 1.88 kg/s (4.145 lb/s) at a pressure of 1.433 MPa (14.15 atm) is given in Equation AA 9.1.3.6.

$$\text{Vol} = 1.88 \left[\frac{(332)\ (8.30)}{1.433}\right] = 3.62\ \frac{m^3}{s} \qquad \text{(AA 9.1.3.6)}$$

The velocity in the demister is, therefore, 0.592 m/s (1.94 ft/s), and for a 5 s residence time, the height would be 2.96 m (9.71 ft). This vessel is stainless steel clad, rated at 2.026 MPa (20 atm) and 600°K (620°F). The weight is 1.00 Mg (1.102 ton) and the cost $490,000.

Forced Draft Water Cooler. The composition of the gas on inlet to the scrubber is shown in Table AA 9.1.3.1

Table AA 9.1.3.1

Scrubber Inlet Gas Composition

Component	Mole
H_2	0.1522
CO	0.2055
CO_2	0.2753
H_2O	0.3068
CH_4	0.0601

and has an enthalpy given by the equation

$$\text{Enthalpy} = 30.851 + 11.553 \times 10^{-3}\ T^2 \text{ kJ/kg mole gas}$$

For an inlet temperature of 541°K (514°F), a flow of 1.88 kg moles/s (4.145 lb moles/s), and a molecular weight of 27.70, not including the entrained S, the enthalpy flux is 23.789 MJ/s (22,552 Btu/s [222°K (-60°F) base temperature]. The outlet composition is shown in Table AA 9.1.3.2.

Table AA 9.1.3.2

Scrubber Outlet Composition

Component	Mole
H_2	0.2167
CO	0.2927
CO_2	0.3921
H_2O	0.0128
CH_4	0.0856

and has an enthalpy given by the equation

$$\text{Enthalpy} = 25.953 + 12.767 \times 10^{-3}\ T^2 \quad \text{kJ/kg mole gas}$$

and a heating value of 220,447 MJ/kg mole (9479 Btu/lb mole). At a gas temperature 332°K (137°F) and a flow rate of 1.32 kg moles/s (2.91 lb moles/s), a molecular weight of 27.50, the enthalpy flux is 4.797 MW (4547 Btu/s). The water temperature is allowed to reach 420°K (296°F). Considering the heat capacity of water to be 4.186 kJ/kg°K (1 Btu/lb°F), the water rate needed is:

$$\text{Water rate} = \frac{(5683-1146)}{1\ (420-332)} = 51.6 \text{ kg/s}$$

From Perry (Reference 9.15) the water tower concentration rate is at least 2.033 kg/m^2 (0.416 lb/s ft^2). Running the tower at 110% saturation to allow for surges in the water rate, the total fan power needed is 11.27 kW. The needed pump power is 210 kW (assuming an efficiency of 50%). The total surface area needed is 24.6 m^2 (264.8 ft^2). If a packed column using Tellerette Packing by Celcate is utilized, which has 181.45 m^2/m^3 (55.3 ft^2/ft^3), weighs 120.15 kg/m^3 (7.5 lb/ft^3), and

costs \$495/m^3 (\$14.02/ft^3) for the 2.54 cm (1 in) size, then 0.272 m^3 (9.606 ft^3) of packing is needed, allowing for an efficiency of 50% in the performance of the packing. The water evaporation rate is 8.4 kg/s (18.51 lb/s). Allowing 2% of the water circulation rate for blowdown, the total makeup water needed is 8.6 kg/s (18.96 lb/s). The tower is 1 m (3.28 ft) in diameter and 2 m (6.56 ft) high, made of stainless steel clad carbon steel. The total weight is 1 Mg (1.101 tons) and it costs \$140,000.

Heat Exchangers and Steam Generators. The composition of the process gas from the heat exchangers is given in Table AA 9.1.3.3

Table AA 9.1.3.3

Heat Exchanger Outlet Gas Composition

Component	Mole Fraction
H_2	0.1522
CO	0.2055
CO_2	0.2753
H_2O	0.2078
H_2S	0.0990
CH_4	0.0601

molecular weight = 26.30

$$\text{Enthalpy} = 27.042 + 10.967 \times 10^{-3}\ T^2 \quad \text{kJ/kg mole gas}$$

at a gas rate of 1.88 kg mole/s (4.145 lb mole/s), and an outlet temperature of 541°K (514°F), the energy flux out due to the gas is 21.236 MW (20132 Btu/s). The energy content of the potassium carbonate and potassium sulfate in the outlet streams is

$$\text{Energy } K_2CO_3 = 125 \frac{\text{kJ}}{\text{kg mole}} (541-222) \left[\frac{32.71}{174 \dfrac{\text{kg}K_2SO_4}{\text{kg mole}}} 0.869 \frac{\text{kg}}{\text{s}} K_2SO_4 \right] = 6.522 \text{ MW}$$

$$\text{Energy } K_2SO_4 = 139 \frac{\text{kJ}}{\text{kg mole}} (541-222) \times \frac{32.71}{174} \times 0.131 = 1088 \text{ kW}$$

The energy flux due to the gasifier gas into the regenerator given by the equation, enthalpy = $27\ T + 7.2 \times 10^{-3}\ T^2$ kJ/kg mole gas, is 69.705 MW

The energy flow of the potassium sulfate inlet stream is

$$139 \frac{\text{kJ}}{\text{kg mole}} (350-222) \frac{32.71}{174} = 3.332 \text{ MW}$$

For the reaction

$$K_2SO_4 + 4H_2 + CO_2 \rightarrow K_2CO_3 + H_2S + 3H_2O$$

the standard free energy of reaction is -64.611 MJ/kg mole potassium sulfate. The energy generated by the reaction is therefore

$$-64.611 \left(\frac{\text{MJ}}{\text{kg mole } K_2SO_4}\right) \left(32.71 \frac{\text{kg}K_2SO_4}{174 \dfrac{\text{kg}K_2SO_4}{\text{kg mole } k_2SO_4}}\right) (0.869) = -10.553 \text{ MW}$$

For the water gas shift reaction

$$CO + H_2O \rightarrow CO_2 + H_2$$

the standard free energy of reaction is -41.1 kJ/kg mole of carbon monoxide. The energy supplied by this reaction would be

$$(-41.198 \frac{\text{MJ}}{\text{kg mole CO used}})(0.3717 \frac{\text{kg mole CO}}{\text{kg mole gas}})(2.04 \frac{\text{kg mole gas}}{\text{s}})(0.50)$$

$$= 15.601 \text{ MW}$$

For an ash composition as given in Table AA 9.1.3.4

Table AA 9.1.3.4

Composition of the Ash

Component	Mole Fraction
SiO_2	0.6788
Al_2O_3	0.1268
Fe_2O_3	0.0871
CaO	0.0921
MgO	0.0151

molecular weight = 71.65

Enthalpy = $55.9\ T + 37.3 \times 10^{-3}\ T^2$ kJ/kg mole

For a flow rate of 3.84 kg/s (8.466 lb/s), the net energy requirement for the ash is

$$55.9\ (541-350)\ \left(\frac{3.84}{71.65}\right) + \left[37.2 \times 10^{-3}\ (541^2 - 350^2)\ \frac{3.84}{71.65}\right] = 218\ \text{kW}$$

A total energy balance around the regenerators and heat exchangers is given in Table AA 9.1.3.5.

Table 9.1.3.5

Regenerator Energy Balance

IN	Gas	69.705	
	K_2SO_4	3.332	
			+73.037
-OUT	Ash	0.913	
	K_2SO_4	1.088	
	K_2CO_3	6.522	
	Gas	21.085	
			-29.608
-Reactions		-26.154	
			-26.154
TOTAL removed by heat exchanger			69.583 MW

The gas temperature going into the heat exchangers is determined by an energy balance around the regenerator.

IN	Gas	69.705	
	K_2SO_4	3.332	
-Reactions		-26.154	
TOTAL Enthalpy IN			99.191 kW

Regenerator must be maintained at around 1040°K (1412°F). The enthalpy of the regenerator gas at 1040°K (1412°F) is 76,574 kW Therefore, 22.617 MW (21441 Btu/s) must be removed. This is accomplished by running the cold postscrubber gas through the outer jacket of the regenerator. The new temperature of the cooling gas is 760°K (908°F) for an inlet temperature of 332°K (137°F), and a composition as shown in Tabel AA 9.1.3.6,

Table 9.1.3.6

Regenerator Cooling Gas Composition

Component	
H_2	0.2167
CO	0.2927
CO_2	0.3921
H_2O	0.0128
CH_4	0.0856

an enthalpy given by $0.84\ T + 12.77 \times 10^{-3}\ T^2$ kJ/kg mole gas, a gas flow rate of 1.32 kg/s (2.91 lb/s), and a hot inlet temperature of 960°K (1268°F). The mean temperature difference is then calculated as

$$\Delta T_{mean} = \frac{(960-760) + (1040-332)}{2} = 454°K$$

From Perry (Reference 9.15) an overall heat transfer coefficient of about $47.3\ W/m^2°K$ (8.33 Btu/hr-ft^2-°F) may be expected. The total area required is, therefore,

$$A = \frac{Q}{U\ \Delta T_{mean}}$$

$$A = \frac{22617}{(1.47)(454)} = 340\ m^2$$

The total cooling area of the regenerators is 340 m^2. This is therefore a sufficient area to provide the required cooling. The warmed up post-Claus gas is then used to cool the regenerator outlet gases after expansion in the turbine. The total enthalpy to be exchanged is

$$69.584\ MW - 22.617\ MW = 46.967\ MW$$

The hot gas inlet temperature is 1040°K (1412°F), and the cold gas [760°K (908°F)] is then sent to the turbine to be expanded. For an adiabatic expansion from 1.418 to 6.07 MPa (14 to 6 atm), the outlet temperature is 598°K (616°F). This gas is then used to cool the hot gas outlet of the regenerator. It is desirable to generate the steam needed for the gasifier at 478°K (400°F) from this hot gas stream or the final cooling step.

The energy required for this is 32.052 MJ/s (30,385 Btu/s). If the final hot gas temperature is 541°K (513°F), this requires that the gas enter the stream generator at 854°K (1078°F). The mean temperature difference in the steam generator is

$$\Delta T_{mean} = \frac{(854 - 478) + (541 - 478)}{2} = 220°K$$

From Perry (Reference 9.15), the expected overall heat transfer coefficient for a shell and tube exchanger is about $72.7\ W/m^2K$ (12.81 Btu/hr-ft^2-°F). The required surface area would then be

$$A = \frac{Q}{U \Delta T_{mean}}$$

$$A = \frac{29089}{(2.26)(220)} = 585 \text{ m}^2$$

Two 300 m^2 stainless steel steam generators are used. Each is 5 m (16.4 ft) long [4.78 m (15.68 ft) by 2.54 cm (1 in) id], with a shell diameter of 1.07 m (3.51 ft). They are rated at 2.026 MPa (20 atm) and 1200°K (1700°F). Each weighs 13.182 Mg (14.42 ton) and costs $270,000.

Finally, the size of the heat exchangers to reduce the hot gas temperature from 1040°K to 854°K (1412 to 1077°F) is calculated. The total enthalpy to be transferred is

$$69.584 - 22.617 - 32.052 = 14.915 \text{ MW}$$

The cold gas leaves at a temperature of 899°K (1158°F). The log mean ΔT is

$$\Delta T_{\ell n\ mean} = \frac{(1040-899) - (854-598)}{\ell n\ (\frac{1040-899}{854-598})} = 193°K$$

The total area required is

$$A = \frac{Q}{U \Delta T_{\ell n\ mean}}$$

$$A = \frac{14915}{(1.47)(193)} = 527 \text{ m}^2$$

Two 275 m^2, stainless steel U-tube exchangers are used. Each is 5 m (16.4 ft) long [4.89 m (16.04 ft) by 2.54 cm (1 in) id], with a shell diameter of 1.07 m (3.51 ft). They are rated at 2.026 MPa and 1200°K (1700°F). Each weighs 15.115 Mg (16.661 ton) and costs $310,000.

Sulfur Settling Tank. As recommended in Perry (Reference 9.15), a 600 s (10 m) holding time will be used. The tank will be horizontal with an impact plate. The tank is rated at 2.026 MPa (20 atm) 500°K (440°F) and will be stainless steel clad. The total required volume is 35 m^3 (1236 ft^3) with a diameter of 2 m (6.56 ft) and a length of 11.1 m (36.42 ft). The cost of the vessel is $180,000 and the weight is 43.727 Mg (48.199 ton). The mass of water rejected with the sulfur is 10.05 kg/s (22.16 lb/s) and the mass of sulfur is 9.05 kg/s (19.95 lb/s).

Sulfur Dioxide Burner. 3.04 kg/s (6.70 lb/s) of the sulfur produced by the settling tank is injected into the sulfur dioxide burner. Here compressed oxygen is used to burn the sulfur to sulfur dioxide. The sulfur is fed to the burner from a lockhopper which is heated by the burner to achieve some degree of drying before injection. The high-pressure burner is a cam type. The size is approximately 1.52 m (4.98 ft) diameter and 2.13 m (6.99 ft) high. The air feed rate is 3.43 kg/s (7.56 lb/s) at 1.722 MPa (17 atm) and 668°K (742°F). The vessel is stainless steel clad and rated at 2.026 MPa (20 atm) and 1200°K (1700°F). The weight is approximately 5.909 Mg (6.51 ton) and costs $160,000.

Sulfur Dioxide Burner Lockhopper. Holdup time is about 1.2 ks (20 min). Two are required. The volume in each is 1.7 m^3 (60.03 ft^3). These vessels are 0.5 m (19.68 in) in diameter and 2.16 m (7.09 ft) high stainless steel clad, and rated at 400°K (260°F) and 2.026 MPa (20 atm). Each weighs 3.865 Mg (4.268 ton) and costs $20,000.

Oxygen Compressors. The air compressors must supply 3.04 kg/s of oxygen to the sulfur dioxide burner and 14.20 kg/s of oxygen to the coal gasifiers. Using the formula (from Reference 9.12)

$$\text{Power, kW} = \frac{6.89}{E} P_{inlet} V_{inlet} \frac{X}{K}$$

where the efficiency, E, is taken as 50%, the P inlet is 103.4 kPa (15 psi) abs, V_{inlet} is 11.935 m^3/s (25,285 scfm), X/K for oxygen is 0.881, the power is found to be 2172 kW. The outlet temperature is (Reference 9.15)

$$T_{out} = T_{in}\ (X+1) = 300\ (2.23) = 669°K$$

The cost of this compressor, which requires cast steel casings and is rated at 2044 kW, is $600,000.

Lock Gas Compressor. The lock gas requirements are 0.218 sm^3/s (463 scfm) for the potassium sulfate locks, 0.0185 sm^3/s (39.2 scfm) for the coal locks, and 0.0015 sm^3/s (3.18 scfm) for the sulfur locks. The gas to be compressed is the heated fuel gas from 1.418 to 1.722 MPa (14 to 17 atm). The power required is 33.8 kW. The outlet temperature is 700°K (800°F). One is required. The cost is $110,000.

Gas Turbine. The gas rate through the turbine is 36.3 kg/s (80 lb/s). The gas is expanded from 1.418 to 0.608 MPa (14 to 6 atm). The power output is 595 kW, and the gas outlet temperature is 598°K (616°F). The cost is about $270,000.

Electric Motor. This motor supplies the 1483 kW of power not supplied by the gas turbine. The cost of this unit is $100,000.

Oxygen Plant. A total of 17.24 kg/s (381 lb/s) of oxygen are needed. At an energy cost of 936 kJ/kg (402.49 Btu/lb) of oxygen, the total power needed is

$$\text{Energy cost} = (936)\ (17.24) = 16136 \text{ kW}$$

The cost of an oxygen plant (Reference 9.21), installed is given by the formula

$$\text{cost, \$} = \$17 \times 10^6 \left(\frac{X}{20 \text{ kg/s } O_2}\right)^{0.8215}$$

where X is the plant capacity in kg/s.
For a 17.24 kg/s plant, the cost is $15,050,000.

Overall mass and energy balances are presented in Tables AA 9.1.3.7 and AA 9.1.3.8, respectively. Table AA 9.1.3.9 presents an energy balance on the gasifier. Table AA 9.1.3.10 breaks down the cost of the plant according to the price of equipment. Figure AA 9.1.3.1 presents the layout for the seed regeneration plant. Figure AA 9.1.3.2 is a detailed flow diagram for this process. Table AA 9.1.3.11 is the accompanying flow chart.

Table AA 9.1.3.7 - Overall Mass Balance on Seed Regeneration System for Base Case Number 2

IN, kg/s	
Coal	21.72
Water for Steam	11.89
Air	73.99
K_2SO_4	32.71
Ash	3.84
Makeup Water	8.6
Dryer Air	3.91
TOTAL IN	156.66
OUT, kg/s	
MHD Gas	35.2
Evaporated & Blowdown Water	8.6
Sulfur and Water	16.13
Dryer Gases	7.18
Ash Slurry	2.09
Ash	3.84
$K_2CO_3 + K_2SO_4$	26.82
Rejected N_2	56.75
TOTAL OUT	156.61

% Error = 0.15/160.41 x 100% = 0.032%

Table AA 9.1.3.8 - Energy Balance on Seed Regeneration System for Base Case Number 2

Energy Rate In			
Coal (21.72 kg/s, 25,036 kJ/kg)			543,849 kW
K_2SO_4 (294°K basis)			1,457
Ash (294°K basis)			234
Power:	Conveyer	7.46 kW	
	Elevator	11.19	
	Ash Pump	15.00	
	Evaporator	221.27	
	Electrostatic Precipitator	9.20	
	Motor	1483.00	
	Dryer	298.00	
	O_2 Plant	16136.00	
		18181.12 kW	
	@	50% eff.	36,381
TOTAL			581,921 kW
Energy Rate Out			
Heating Value of MHD Gas (HHV=220,447 kJ/kg)			282,174 kW
Enthalpy			31,893
Compression Energy (1407 kW at 50% eff.)			2,817
			316,884 kW
NET ENERGY USAGE			265,037 kW
NET COAL USAGE			10.5 kW

Table AA 9.1.3.9 - Energy Balance on Gasifier for Base Case Number 2

IN (kW)		
Steam (enthalpy and compression)	32,052	
O_2 (enthalpy and compression)	9,234	
Coal (HHV)	543,849	
TOTAL IN	585,136	kW
OUT (kW)		
Enthalpy	69,705	
HV	477,589	
Compressor	4,161	
TOTAL OUT	551,455	kW
EFFICIENCY OF GASIFIER	94.2%	

Table AA 9.1.3.10 – Installed Costs for Seed Regeneration Equipment for Base Case Number 2

Item	Number	Cost Each	Total Cost
K_2SO_4 Surge Bin	3	$ 120,000	$ 360,000
K_2SO_4 Locks	3	80,000	240,000
Coal Bin	1	50,000	50,000
Coal Reclaimer Conveyer	1	100,000	100,000
Coal Crusher Surge Bin	1	90,000	90,000
Coal Crusher/Dryer	1	1,280,000	1,280,000
Coal Surge Bins	2	210,000	420,000
Predried Coal Elevator	1	70,000	70,000
Sized Coal Feed Locks	2	60,000	120,000
Sized Coal Feed Hoppers	2	60,000	120,000
Coal Preheater	1	80,000	80,000
Volatilizers	2	680,000	1,360,000
Gasifiers	2	520,000	1,040,000
Ash Quench Pots	2	60,000	120,000
Ash Slurry Locks	2	60,000	120,000
Ash Slurry Pumps	2	10,000	20,000
Regenerators	2	1,490,000	2,980,000
Ash Cyclones	1	240,000	240,000
K_2CO_3 Electrostatic Precipitators	1	1,040,000	1,040,000
Ash Lockhoppers	2	160,000	320,000
K_2CO_3 Lockhoppers	2	420,000	840,000
Claus Reactor	1	210,000	210,000
Scrubber/Demister	1	490,000	490,000
Cooling Tower	1	140,000	140,000
Steam Generators	2	270,000	540,000
Heat Exchangers	2	310,000	620,000

(continued)

Table 9.1.3.10 (Continued) - Installed Costs for Seed Regeneration Equipment for Base Case Number 2

Item	Number	Cost Each	Total Cost
Settling Tanks	1	$ 180,000	$ 180,000
SO_2 Burner	1	160,000	160,000
SO_2 Lockhopper	2	20,000	40,000
O_2 Compressor	1	600,000	600,000
Lock Gas Compressor	1	110,000	110,000
Gas Turbine	1	270,000	270,000
Electric Motor	1	90,000	90,000
Oxygen Plant	1	15,050,000	15,050,000
TOTAL			$29,510,000

Dwg. 257C230

TABLE AA 9.1.3.11 – SEED REGENERATION FLOW CHART FOR BASE CASE 2

Flow Name	Coal	Power	Coal	Power	Air	Exhaust	Dry Coal	Power	Hot Coal	Power	Ash	Syn. Gas	O_2	Steam	Lock Gas
Flow Number	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
Mass Rate, kg/s	21.72		21.72		3.91	7.18	19.54		19.54		2.09	43.5	14.20	11.89	0
Temperature, °K	300		300		300	370	370		1000		1144	1144	669	478	802
Pressure, kPa	101		101		101	101	101		1520		1520	1520	1722	1722	1722
Power, kW		7.46		298				11.19		15					
N_2, Mole Fraction					.79	.7066						0	0	0	0
H_2					0	0						.2788	0	0	.2167
CO					0	0						.3764	0	0	.2927
CO_2					0	.1632						.1464	0	0	.3921
H_2O					0	.1048						.1324	0	1.0	.0128
H_2S					0	0						.011	0	0	0
CH_4					0	0						.0551	0	0	.0856
SO_2					0	0						0	0	0	0
O_2					.21	.0254						0	1.0	0	0
S, kg/s															
K_2SO_4															
K_2CO_3															
Ash											2.09				

Flow Name	Fuel Gas	Ash	Seed + Gas	Coal Gas	Med. Gas	Hot Gas	Gas+Solid	Ash + K_2SO_4	Power	Make-up	Evap. Water	H_2O+S	Cool H_2O	Hot H_2O	O_2
Flow Number	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30
Mass Rate, kg/s	1.1	8.05	72.00	80.05	80.05	80.05	80.05	36.55		8.6	8.6	16.13	60.2	60.2	3.04
Temperature, °K	899	541	541	541	854	1040	960	350		300	373	420	332	420	669
Pressure, kPa	607	1418	1458	1469	1484	1499	1520	101		101	101	1418	1722	1722	1722
Power, kW									221						
N_2, Mole Fraction	0		0	0	0	0	0			0	0	0	0	0	0
H_2	.2167		.1522	.1522	.1522	.1522	.2788			0	0	0	0	0	0
CO	.2927		.2055	.2055	.2055	.2055	.3764			0	0	0	0	0	0
CO_2	.3921		.2753	.2753	.2753	.2753	.1464			0	0	0	0	0	0
H_2O	.0128		.2078	.2078	.2078	.2078	.1324			1.0	1.0	1.0	1.0	1.0	0
H_2S	0		.0990	.0990	.0990	.0990	.011			0	0	0	0	0	0
CH_4	.0856		.0601	.0601	.0601	.0601	.0551			0	0	0	0	0	0
SO_2	0		0	0	0	0	0			0	0	0	0	0	0
O_2	0		0	0	0	0	0			0	0	0	0	0	1.0
S, kg/s		0	0	0	0	0	0	0				6.01			
K_2SO_4		.673	3.61	4.28	4.28	4.28	32.71	32.71				0			
K_2CO_3		3.54	19.0	22.54	22.54	22.54	0	0				0			
Ash		3.84	0	3.84	3.84	3.84	3.84	3.84				0			

TABLE AA 9.1.3.11 – SEED REGENERATION FLOW CHART FOR BASE CASE 2 (cont'd)

Dwg. 2570229

Stream Name	SO_2	S	Lock Gas	$S+H_2O$	Cold Gas	S+Gas	Gas	Seed	Power	Lock Gas	Lock Gas	Gas	MHD Gas	Water	Warm Gas
Stream Number	31	32	33	34	35	36	37	38	39	40	41	42	43	44	45
Mass Rate, kg/s	6.08	3.04	0	79.37	36.3	55.47	49.39	22.61		0	0	0	35.2	11.89	36.3
Temperature, °K	541	420	802	420	332	541	541	541		802	802	760	899	300	760
Pressure, kPa	1722	1418	1722	1418	1428	1433	1444	1418		1722	1722	607	607	101	1418
Power, kW									9.2						
N_2, Mole Fraction	0		0	0	0	0	0			0	0	0	0	0	0
H_2	0		.2167	0	.2167	.1522	.1522			.2167	.2167	2167	.2167	0	.2167
CO	0		.2927	0	.2927	.2055	.2055			.2927	.2927	.2927	.2927	0	.2927
CO_2	0		.3921	0	.3921	.2753	.2753			.3921	.3921	.3921	.3921	0	.3921
H_2O	0		.0128	1.0	.0128	.3068	.2078			.0128	.0128	.0128	.0128	1.0	.0128
H_2S	0		0	0	0	0	.0990			0	0	0	0	0	0
CH_4	0		.0856	0	.0856	.0601	.0601			.0856	.0856	.0856	.0856	0	.0856
SO_2	1.0		0	0	0	0	0			0	0	0	0	0	0
O_2	0		0	0	0	0	0			0	0	0	0	0	0
S, kg/s		3.04		9.05		9.05		0							
K_2SO_4		0		0		0		3.61							
K_2CO_3		0		0		0		19.0							
Ash		0		0		0		0							

Stream Name	Expanded Gas	O_2	O_2	Power	Air	N_2	Power	Hot Gas
Stream Number	46	47	48	49	50	51	52	53
Mass Rate, kg/s	36.3	17.24	17.24		73.99	56.75		36.3
Temperature, °K	598	669	300		300	300		899
Pressure, kPa	607	1722	101		101	101		607
Power, kW				1483			16136	
N_2, Mole Fraction	0	0	0		.79	1.0		0
H_2	.2167	0	0		0	0		.2167
CO	.2927	0	0		0	0		.2927
CO_2	.3921	0	0		0	0		.3921
H_2O	.0128	0	0		0	0		.0128
H_2S	0	0	0		0	0		0
CH_4	.0856	0	0		0	0		.0856
SO_2	0	0	0		0	0		0
O_2	0	1.0	1.0		.21	0		0
S, kg/s								
K_2SO_4								
K_2CO_3								
Ash								

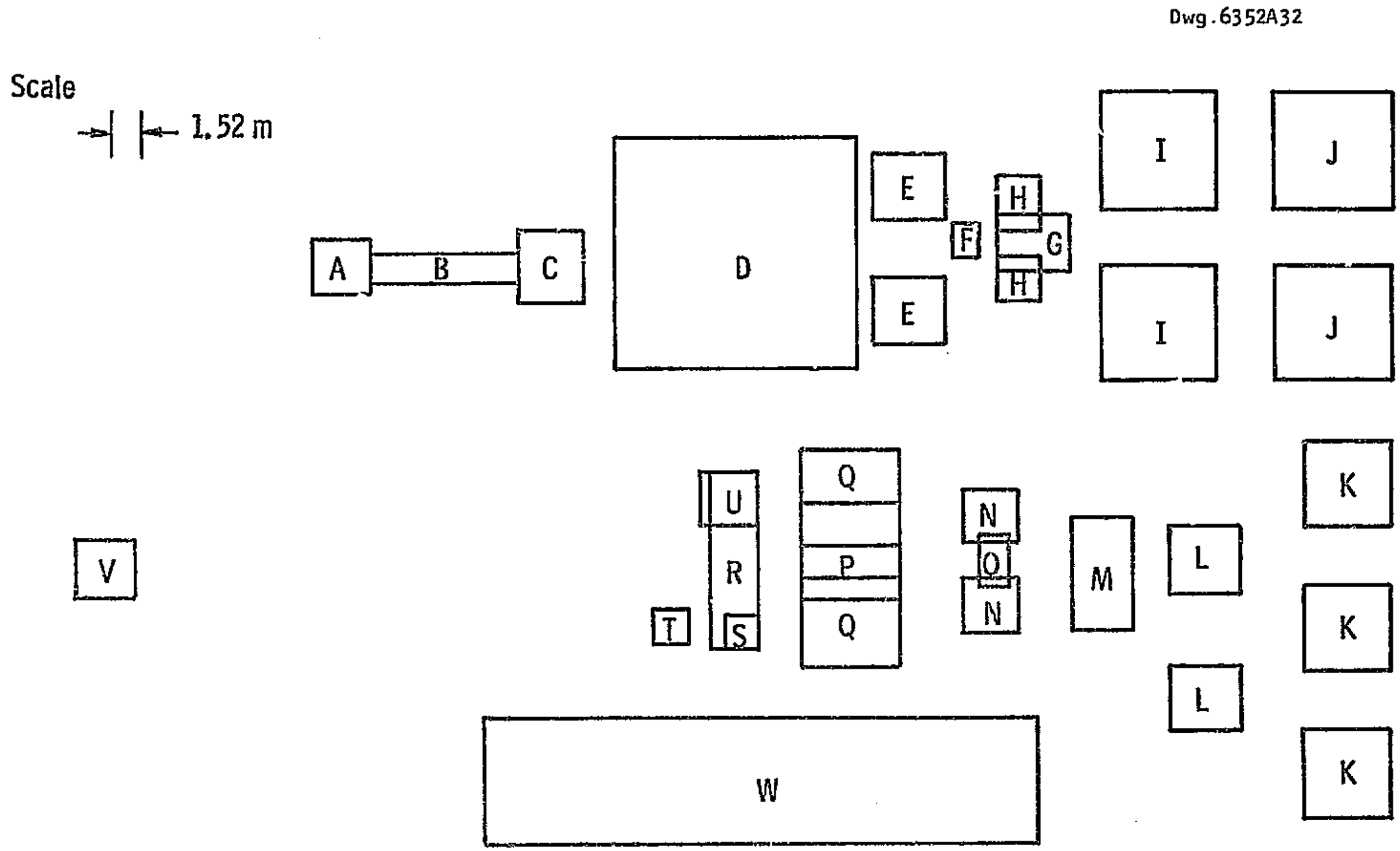


Fig. AA 9.1.3.1—Seed regenerative layout for Base Case 2

Figure AA 9.1.3.1 - Legend for layout diagram and flow sheet
Addendum

A Coal Bin Receiver
B Coal Reclaimer Elevator
C Coal Crusher Surge Bin
D Coal Crusher/Dryer
E Coal Surge Bins
F Predried Coal Elevator
G Coal Preheater
H Sized Coal Feed Lockhopper
I Volatilizers
J Gasifiers
K K_2SO_4 Surge Bins/Lockhoppers
L Regenerators
M Heat Exchangers/Steam Generators
N Ash Lockhoppers
O Ash Cyclone
P K_2CO_3 + K_2SO_4 Electrostatic Precipitator
Q K_2CO_3 + K_2SO_4 Hoppers
R Settling Tank
S Claus Reactor
T SO_2 Burner
U Scrubber/Demister
V Cooling Tower
W O_2 Plant/Turbine-Compressor-Motor Complex

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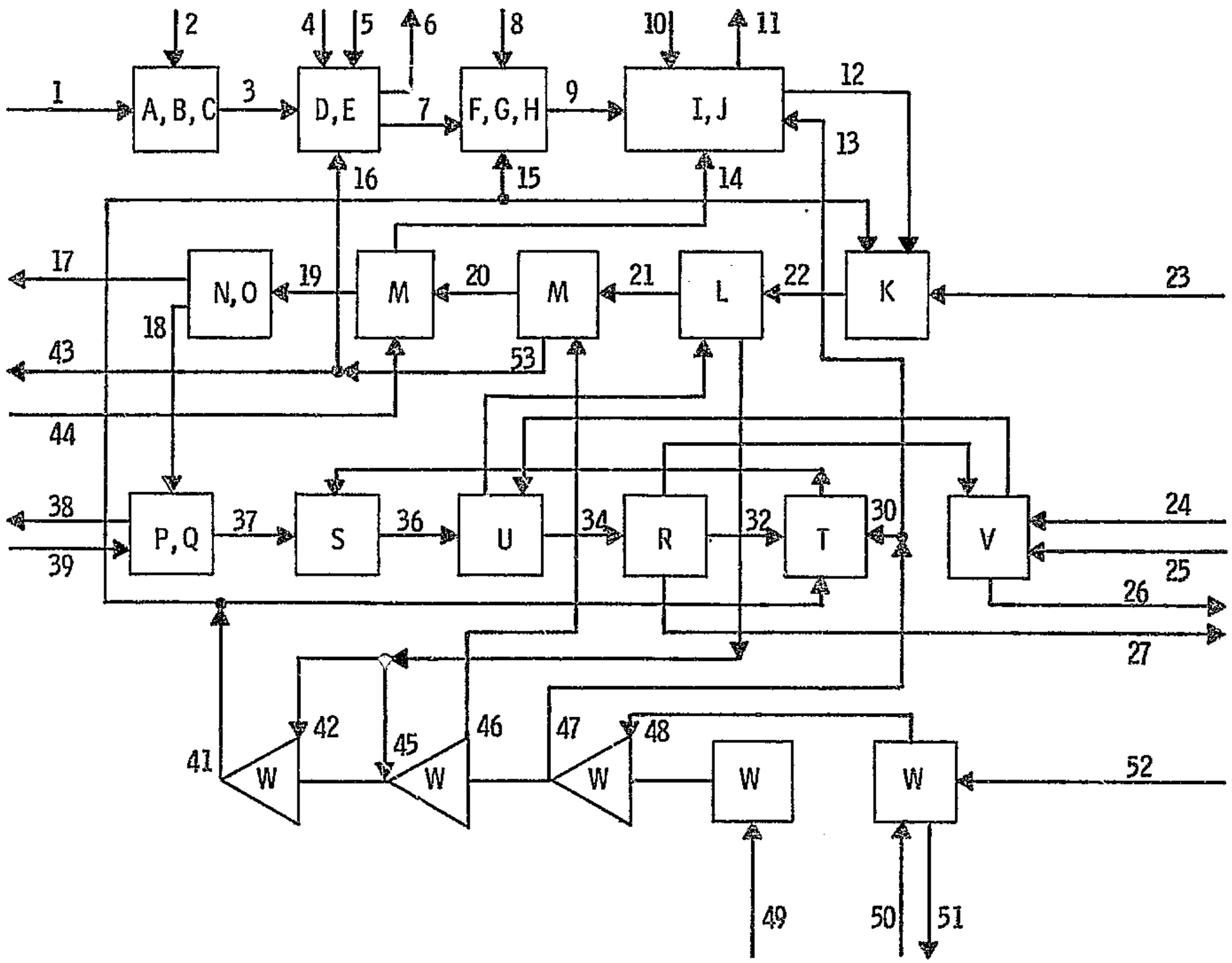

Fig. AA 9.1.3.2—Seed regenerative system flow diagram for Base Case 2

Subappendix AA 9.1.4
LEACHING SYSTEM

The amount of water to be added to the ash and potassium carbonate and potassium sulfate flow is determined by the amount of potassium sulfate in the system for conversions, Y_c (less than 0.75), and by the amount of potassium carbonate in the system for conversions (greater than 0.75). This is demonstrated by equating Equations AA 9.1.4.1 and AA 9.1.4.2 in Equation AA 9.1.4.3.

$$\frac{kgH_2O}{s} = X_{ash,14}\ S_{14}\ \left(\frac{0.6}{94}\right)\left[\frac{174(1-Y_c)}{0.24\ \frac{kgK_2SO_4}{kgH_2O}}\right] = 4.63\ X_{ash,14}\ S_{14}\ (1-Y_c) \qquad \text{(AA 9.1.4.1)}$$

or

$$\frac{kgH_2O}{s} = X_{ash,14}\ S_{14}\ \left(\frac{0.6}{94}\right)\left[\frac{0.138\ Y_c}{\frac{0.567\ kgK_2CO_3}{kgH_2O}}\right] = 1.55\ X_{ash,14}\ S_{14}\ Y_c \qquad \text{(AA 9.1.4.2)}$$

where the crossover point between the above two equations is determined by the equation

$$1.55\ X_{ash,14}\ S_{14}\ Y_c = 4.63\ X_{ash,14}\ S_{14}\ (1-Y_c) \qquad \text{(AA 9.1.4.3)}$$

or

first equation when $Y_c \leq 0.75$
second equation when $Y_c > 0.75$.

The temperature of the water - potassium sulfate-potassium carbonate solution going into the flash drum is now determined.

$$139 \text{ kJ/kgmole } K_2SO_4 \ (X_{ash,14}\ S_{14})(\frac{0.6}{94})(1-Y_c)(541-T) +$$

$$125 \text{ kJ/kgmole } K_2CO_3 \ (X_{ash,14}\ S_{14})(\frac{0.6}{94})(541-T)\ Y_c$$

$$+ 55.92 \text{ kJ/kgmole Ash} \left[\frac{S_{14}\ X_{ash,14}}{71.65}\right] (541-T)$$

$$= (T-373)\ (X_{ash,14}\ S_{14}) \left[\frac{138\ Y_c}{0.567}\right] (\frac{0.6}{94})$$

(since most conversions are greater than 0.75).

Solving for T, one obtains

$$T = 541 \left[\frac{73.3 + Y_c}{97.5 + Y_c}\right] {}^\circ K$$

At a flash drum pressure of 101.3 kPa (1 atm), the temperature is 373°K (212°F) and the enthalpy for evaporation of water is 2260 kJ/kg (972 Btu/lb). The amount of steam evaporated is now determined.

$$1.55\ Y_c\ (X_{ash,14}\ S_{14})\ (T-373) = (\text{steam produced})(540)$$

Solving for the steam produced (and substituting for T) one obtains

$$\text{Steam produced} = 1.55\ Y_c\ X_{ash,14}\ S_{14} \left[\frac{73.3 - Y_c}{97.5 - Y_c} - 0.689\right]$$

Assuming that the steam is supplied to the system at a pressure of 1722 kPa and is saturated, the energy released on condensation is 2377 kJ/kg (1022 Btu/lb). The amount of steam needed is then

$$= 540\ (1.55\ X_{ash,14}\ S_{14}\ Y_c - \text{steam produced})/568$$

or

$$0.872\ X_{ash,14}\ S_{14}\ Y_c \left[1 - 0.592 \quad [\frac{73.3 - Y_c}{97.5 - Y_c}]\right] \text{ kg/s}$$

Assuming cooling water is available at 332°K, and that a 20°K use is allowed before it is returned, the amount of cooling water needed to condense the steam produced and the evaporate from the evaporator is:

$$\frac{1.55\ Y_c\ X_{ash,14}\ S_{14}\ 540}{20}$$

or

$$41.9\ Y_c\ X_{ash,14}\ S_{14}\ \text{kg/s}$$

Using Guthrie (Reference 9.12), the power rating of the pump and its cost was determined. The pump power required is:

$$4.58\ Y_c\ X_{ash,14}\ S_{14}\ \text{kW}$$

and the cost, for a stainless steel in-line pump capable of producing a head of 2026 kPa (20 atm) is:

$$\$2253\ Y_c\ X_{ash,14}\ S_{14}$$

The size and cost of a scrapped film evaporator is obtained next. The mean temperature difference is:

$$\Delta T_{mean} = \frac{(478-373) + (478-373)}{2} = 105°\text{K}$$

Heat transfer coefficients are around 0.407 kJ/s-m^2-°K (0.0833 Btu/s-ft^2-°F). The area of the evaporator is, therefore:

$$A = \frac{\text{water to be evaporated} \times 540}{0.407 \times 105}$$

$$= \left[1.69 - \frac{73.3-Y_c}{97.5-Y_c}\right] Y_c\ X_{ash,14}\ S_{14}\ 19.59\ \text{m}^2$$

From Guthrie, the cost is found to be:

$$\$44{,}774 \left[Y_c\ X_{ash,14}\ S_{14} \left|1.69 - \frac{73.3-Y_c}{97.5-Y_c}\right]\right]^{0.55}$$

The size and cost of a shell and tube condenser is now found. The log mean temperature difference is

$$\Delta T_{\text{log mean}} = \frac{(373-352) - (373-332)}{\ln\left[\frac{(373-352)}{(373-332)}\right]} = 30°K$$

For this type of service, heat transfer coefficients of 0.85 W/m^2-°K (0.152 Btu/hr-ft^2-°F) are obtained (Reference 9.15). The required area is therefore

$$A = \frac{(\text{Steam to be condensed})(540)}{(30°K)(0.203)}$$

or

$$137.4\ Y_c\ X_{ash,14}\ S_{14}\ m^2$$

From Guthrie, the cost is determined to be

$$\$17{,}400\ (Y_c\ X_{ash,14}\ S_{14})^{0.65}$$

The mixing tank is assumed to be a cylindrical vessel with a volume of 1 m^3, (35.31 ft^3), containing baffles to aid in settling the ash from the solution. The vessel is rated at 600°K and 2026 kPa (20 atm) and costs $58,666. Clad stainless steel is used. The flash drum is assumed to be the same size as the mixing tank. However, it is only rated at 400°K (261°F) and 101 kPa (1 atm). It is also stainless steel clad. The cost is estimated to be $40,495.

Figure AA 9.1.4.1 presents a flow sheet of the ash leach system and Table AA 9.1.4.1 presents a flow chart of the ash leach system.

<u>Legend for Ash Leach System</u>

A Settling Tank

B Pump

C Flash Drum

D Evaporator (Scrapped Film)

E Condenser

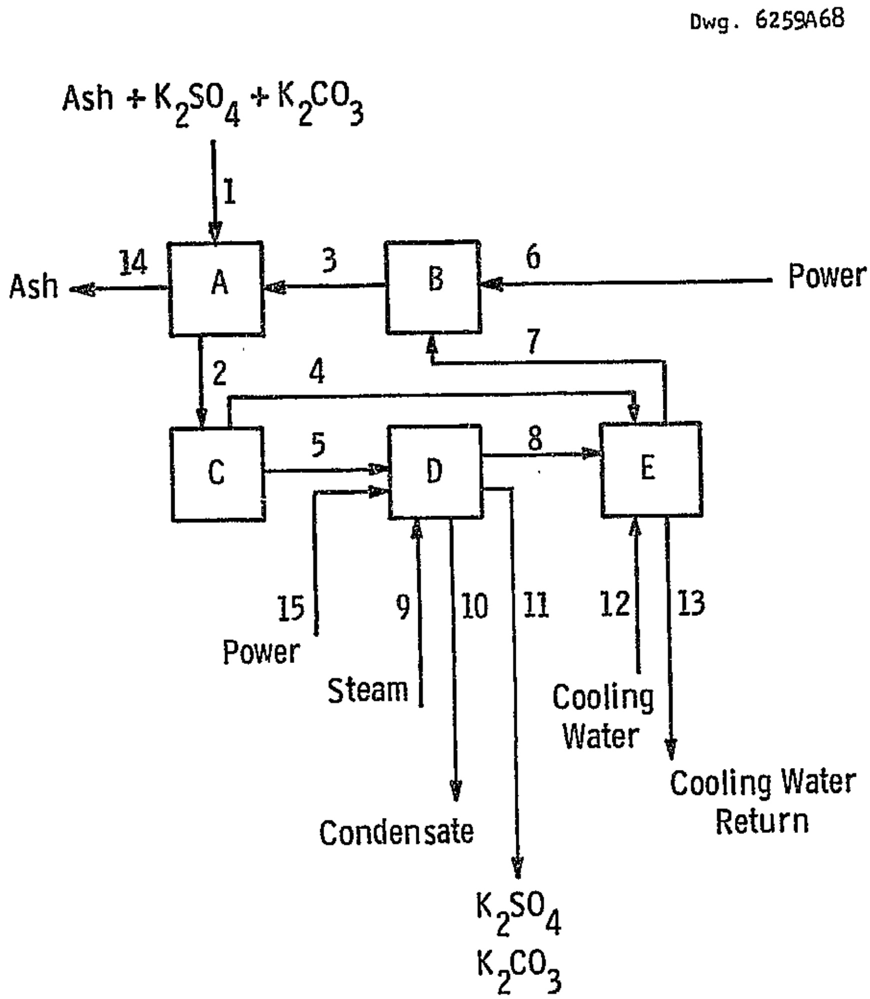

Fig. AA 9.1.4.1—Flow diagram of ash leach system

TABLE AA 9.1.4.1
LEACHING SYSTEM

Stream Name	Feed In	No Ash	Water	Flash	Bottoms	Power	Sat. H_2O	Distillate	Steam	Condensate	Seed	Cooling H_2O	Hot H_2O	Ash	Power
Stream #	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
Mass Rate (kg/s)	A	C	C-B	E	C-E	-	C-B	C-B-E	G	G	B	H	H	A-B	-
Temperature (°K)	541	D	373	373	373	-	373	373	478	373	373	332	352	D	-
Pressure (kPa)	1418	1418	1520	101	101	-	101	101	1722	101	101	101	101	1418	-
Power (kW)	-	-	-	-	-	F	-	-	-	-	-	-	-	-	2
H_2O (kg/s)	0	C-B	C-B	E	C-B-E	-	C-B	C-B-E	G	G	0	H	H	0	-
K_2SO_4+K_2CO_3 (kg/s)	B	B	0	0	B	-	0	0	0	0	B	0	0	0	-
Ash (kg/s)	X	0	0	0	0	-	0	0	0	0	0	0	0	A-B	-

X = Mass rate ash (kg/s) $(X_{ash,14}S_{14})$

f = Fractional conversion, $K_2SO_4 \rightarrow K_2CO_3$ (Yc)

A: $X(1 + f(0.6)(\frac{138}{94}) + (1-f)(0.6)(\frac{174}{94}) = X(2.11 - 0.23f)$

B: $X(1.11 - 0.23f)$

C: $X(1.11 + 1.23f)$

D: $541 (73.3 - f) / (97.5 - f)$

E: $Xf [(73.3 - f) / (97.5 + Yc) - 0.689]$

F: $4.58 \times f$

G: $0.872 \times f [1 - 0.592 (73.3 - f)/(97.5 - f)]$

H: $41.9 \times f$

Subappendix AA 9.1.5

MODIFICATIONS TO SUBAPPENDIX AA 9.1.2
FOR BASE CASE 1

In Base Case 1, the coal is not burned directly in the combustor. Rather, it is first heated to remove most of the volatiles and to produce a char which is then burned in the combustor. This char is considered to retain all the ash and sulfur present initially in the coal. Therefore, S_1 in Subappendix AA 9.1.2 must be modified to exclude the volatile materials removed.

First, the initial coal rate is determined from the power output (MW), heat rate (kJ/kWh), and higher heating value of the coal (kJ/kg) as follows:

$$\frac{\text{(power output)(1000)(heat rate)}}{\text{(hhv)(3600)}} = \frac{\text{kg coal}}{\text{s}}$$

Given the ash-free char rate, the new value for S_1 is obtained as:

$$S_1'' = \left(\text{char rate} + \frac{\text{kg coal}}{\text{s}}\right)\left(\text{\% ash in coal}\right)$$

A new ash fraction is now determined as:

$$X_{ash,1}' = \frac{\left(\frac{\text{kg coal}}{\text{s}}\right)\left(\text{\% ash in coal}\right)\left(100\right)}{\left(\text{Char rate} + \frac{\text{kg coal}}{\text{s}}\right)\left(\text{\% ash in coal}\right)}$$

and a new sulfur fraction, $X_{s,1}'$, is determined as:

$$X_{s,1}' = \frac{\left(\frac{\text{kg coal}}{\text{s}}\right)\left(\text{\% sulfur in coal}\right)\left(100\right)}{\left(\text{Char rate} + \frac{\text{kg coal}}{\text{s}}\right)\left(\text{\% ash in coal}\right)}$$

With these modifications, the material balance equations derived in Subappendix AA 9.1.2 may now be used.

Subappendix AA 9.1.6
SCALING FACTOR FOR PROCESS VESSELS

To determine the scaling factor to use to scale the process vessels for different flow rates, the following procedure was used. Various diameter/height ratios were used to determine, as a function of volume, the base cost of process vessels. Table AA 9.1.6.1 presents the various calculations and Figure AA 9.1.6.1 presents the graph of the Log_{10} Cost versus Log_{10} Volume.

The scaling factor arrived at was 0.666. The data was obtained from Guthrie (Reference 9.12). This scaling factor can be used for those parts of the system which are all pressure vessels or nearly so, and whose size would be proportional to the volume handled. Those vessels which handle the process gas and/or are part of the gasification system will have a size proportional to the amount of sulfur removed (or the flow rate of potassium sulfate into the regeneration system times the percent regeneration). Those vessels handling either ash or potassium sulfate or both will have sizes proportional to the flow rate of potassium sulfate and/or ash into the regeneration system. Table AA 9.1.6.2 presents a list of equipment along with the factors assumed to affect their size.

The total cost of pressure vessel and gasifier equipment affected by the fractional conversion and the potassium sulfate rate is $9 million. Other equipment items affected by the potassium sulfate rate and the fractional conversion include turbine-compressors (scaling factor 0.59) $870,000, motors and mechanical equipment (scaling factor 1) $1,500,000, steam generators and heat exchangers (scaling factor 0.71) $1,160,000, and the oxygen plant (scaling factor 0.8215) $15,050,000. Pressure vessel and compressor costs affected by the potassium sulfate and ash rate are $600,000 and $110,000 respectively. Pressure vessel

Curve 680300-A

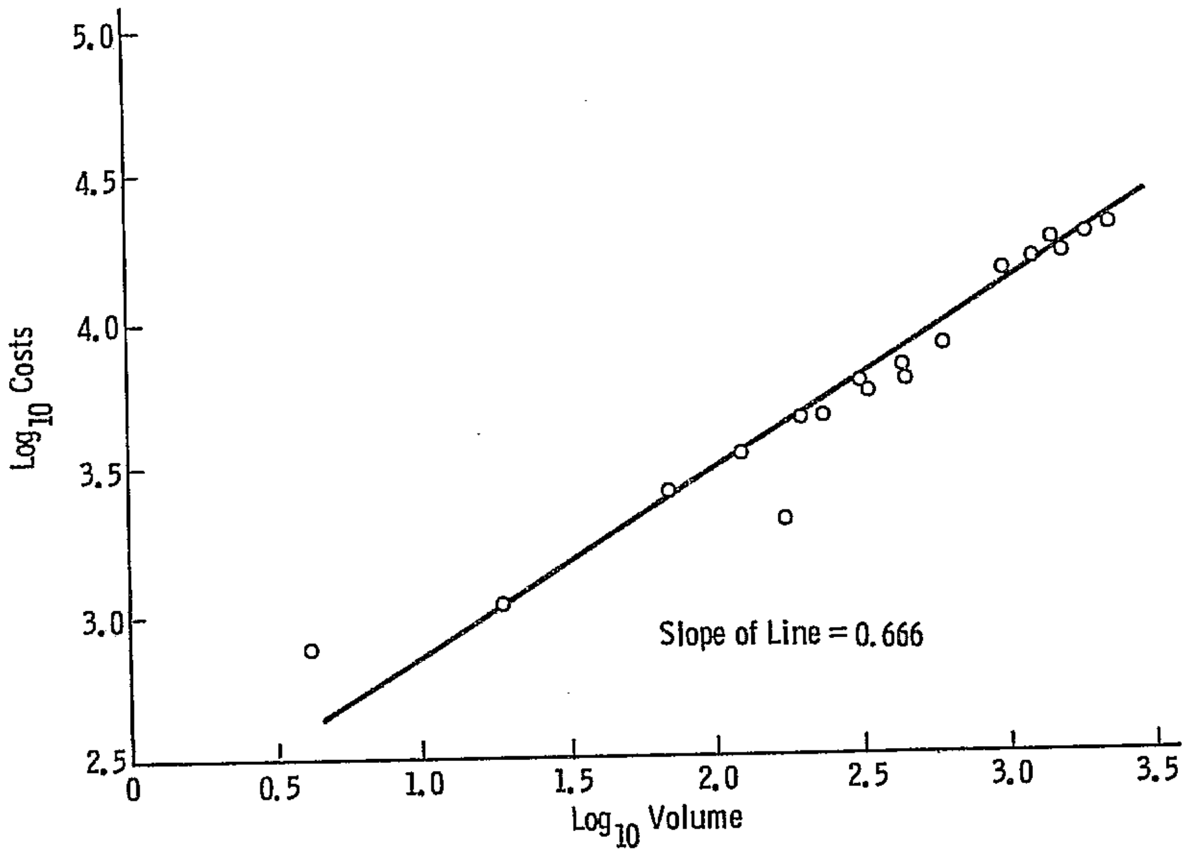

Fig. AA 9.1.6.1—Volume versus cost for variously sized vessels

Table AA 9.1.6.1 - Volume Versus Cost for Variously Sized Vessels

Diameter, ft	Height, ft	Log Volume	Log Cost
2	4	0.602	2.88
3	8	2.25	3.30
4	16	2.30	3.65
5	16	2.50	3.78
6	16	2.66	3.83
7	16	2.80	3.90
8	20	3.00	4.15
9	20	3.10	4.18
10	20	3.20	4.20
2	6	1.28	3.02
3	10	1.85	3.40
4	10	2.10	3.54
5	12	2.37	3.65
6	12	2.53	3.74
7	12	2.66	3.78
8	30	3.18	4.24
9	30	3.28	4.28
10	30	3.37	4.30

Table AA 9.1.6.2 - Equipment List

Item	Fractional	Ash Rate	K_2SO_4 Rate
K_2SO_4 Surge Bin	no	yes	yes
K_2SO_4 Locks	no	yes	yes
Coal Bin	yes	no	yes
Coal Reclaimer Con.	yes	no	yes
Coal Crusher Surge Bin	yes	no	yes
Coal Crusher/Dryer	yes	no	yes
Coal Surge Bins	yes	no	yes
Predried Coal Elevator	yes	no	yes
Sized Coal Feed Locks	yes	no	yes
Sized Coal Feed Hoppers	yes	no	yes
Coal Preheater	yes	no	yes
Volatilizers	yes	no	yes
Gasifiers	yes	no	yes
Ash Quench Pots	yes	no	yes
Ash Slurry Locks	yes	no	yes
Ash Slurry Pumps	yes	no	yes
Regenerators	yes	no	yes
Ash Cyclones	yes	no	yes
K_2CO_3 Electrostatic Precipitator	yes	no	yes
Ash Lockhoppers	no	yes	no
K_2CO_3 Lockhoppers	no	no	yes
Claus Reactor	yes	no	yes
Scrubber/Demister	yes	no	yes
Cooling Tower	yes	no	yes
Heat Exchanger	yes	no	yes
Settling Tank	yes	no	yes
SO_2 Burner	yes	no	yes

(con't)

Table AA 9.1.6.2 (continued)

Item	Fractional	Ash Rate	K_2SO_4 Rate
SO_2 Burner Lockhopper	yes	no	yes
Air Compressors	yes	no	yes
Lock Gas Compressors	no	yes	yes
Gas Turbine	yes	no	yes
Electric Motor	yes	no	yes
O_2 Plant	yes	no	yes
Steam Generator	yes	no	yes

costs affected only by the ash rate total $320,000 and those affected only by the potassium sulfate rate total $840,000. All scaling factors were determined from Reference 12.

For this case, the potassium sulfate rate was 32.71 kg/s, (72.0 lb/s), the regeneration rate 0.869, and the ash rate 3.84 kg/s (8.45 lb/s). These factors can then be written as:

$$\text{Cost of Pressure Vessel, Potassium Sulfate and Fractional Conversion, \$} = 966{,}509\ [(K_2SO_4)(F_K)]^{0.666}$$

$$\text{Cost of Turbine-compressors, \$} = 120{,}735\ [(K_2SO_4)(F_K)]^{0.59}$$

$$\text{Cost of Motors and Mechanical, \$} = 54{,}881\ [(K_2SO_4)(F_K)]$$

$$\text{Cost of Steam Generators and Heat Exchangers, \$} \quad 107{,}729\ [(K_2SO_4)(F_K)]^{0.71}$$

$$\text{Cost of Oxygen Plant, \$} = 962{,}000\ [(K_2SO_4)(F_K)]^{0.82}$$

$$\text{Cost of Pressure Vessels, Ash and Potassium Sulfate Rate, \$} = 54{,}492\ [(K_2SO_4) + \text{Ash}]^{0.666}$$

$$\text{Cost of Compressors, \$} = 13{,}161\ [(K_2SO_4) + \text{Ash}]^{0.59}$$

$$\text{Cost of Pressure Vessels, Ash Rate, \$} = 130{,}507\ (\text{Ash})^{0.666}$$

$$\text{Cost of Pressure Vessels, Potassium Sulfate, \$} = 82{,}147\ (K_2SO_4)^{0.666}$$

The cost of adding a leaching system is:

$$\$99{,}161 \left[1 + F_K(\text{Ash}) \left\{0.0227 + 0.45 \left[1.69 - \left(\frac{73.3 - F_K}{97.5 - F_K}\right)\right]^{0.55}\right\}\right.$$

$$\left.0.175 \left[\left(F_K\right)\left(\text{Ash}\right)\right]^{0.65}\right]$$

where F_K is the fractional conversion and Ash is the ash flow rate.

Utilities usage is directly proportional to the fractional conversion times the potassium sulfate flow rate. For a coal rate of 21.72 kg/s (47.8 lb/s), a water makeup rate of 18.13 kg/s (42.1 lb/s), and a power of 17,405 kW, the utilities factors are:

$$0.612(F_K)(K_2SO_4) \text{ kW}$$

$$0.673(F_K)(K_2SO_4) \text{ kg/s of water}$$

$$0.764(F_K)(K_2SO_4) \text{ kg/s of coal}$$

The coal rate must be multiplied by the factors 0.925 for Montana coal and 1.03 for North Dakota coal. These factors were determined from data presented by Hamm (Reference 9.14). The addition of a leaching system requires the following utilities to be added:

$$\text{Steam Required, kg/s} = 0.872(F_K)(\text{Ash}) \left[1 - 0.592 \left(\frac{73.3 - F_K}{97.5 - F_K}\right)\right]$$

Power Required for Leaching System, kW = $2 + 4.58(F_K)(\text{Ash})$

Water Required, kg/s = $41.9(F_K)(\text{Ash})$

Return utilities include MHD gas, sulfur and water slurry, ash, and potassium sulfate and potassium carbonate:

- MHD gas at 899°K (1159°F) and 608 kPa (6 atm) with a net energy flow rate:

  MHD gas, kW = $11{,}147(F_K)(K_2SO_4)$

- Sulfur and water flow rate at 541°K (514°F) and 101 kPa (1 atm):

  Sulfur, kg/s = $0.212(F_K)(K_2SO_4)$

  Water, kg/s = $0.418(F_K)(K_2SO_4)$

- Ash at 541°K (514°F), potassium sulfate, and potassium carbonate flow rates to the first stage of the combustor:

  Ash, kg/s = $\text{Ash}(2.11 - 0.229\ F_K)$

- Potassium sulfate and potassium carbonate at 541°K (514°F) to the second stage of the combustor:

  Seed, kg/s = $(K_2SO_4)(1 - 0.207\ F_K) - 1.11\ \text{Ash}$

For Montana and North Dakota coals, the MHD gas energy rate must be multiplied by 0.6 and 0.413, respectively (determined from data presented by Hamm, Reference 9.14). If a leaching system is added, the return utilities include:

- Ash at 400°K (261°F) (to disposal).

- Potassium sulfate and potassium carbonate at 541°K (514°F) to second stage of the combustor, kg/s =

$$(1 - 0.207\ F_K)(K_2SO_4)$$

- Cooling water at 353°K (176°F), kg/s =

$$41.9(F_K)(Ash)$$

- Condensed steam at 400°K (261°F) and 1722 kPa (17 atm), kg/s =

$$0.872(F_K)(Ash)\left[1 - 0.592\left(\frac{73.3 - F_K}{97.5 - F_K}\right)\right]$$

Subappendix AA 9.1.7

CALCULATION OF THE MINIMUM EFFICIENCY REQUIRED FOR THE ELECTROSTATIC PRECIPITATOR IN BASE CASE 3

The amount of particulate emissions allowed up the stack under current Federal regulations is 0.043 kg/GJ (0.1 lb/10^6 Btu) (Reference 9.14). Therefore, the total particulate allowed up the stack is:

Stack Particulate, kg/s =

$$\frac{(1000)(\text{power output, MW})(\text{heat rate, kJ/kWh})(0.043\ \frac{\text{kg particulate}}{\text{GJ}})}{3600\ \text{s/hr}}$$

In Base Case 3, the only solids in the MHD duct are the pollucite (Cs Al Si_2O_6) used to make the MHD gases conductive. Given the seed rate (in terms of cesium carbonate), then the amount of pollucite in the gas stream (in a carbonated form) is:

$$\text{Pollucite Stack, kg/s} = (\frac{\text{seed rate}}{326})(668)$$

The fraction of particulates allowed up the stack is, therefore:

$$Y_e = \frac{(1000)(\text{power output, MW})(\text{heat rate, kJ/kWh})(\frac{0.043\ \text{kg}}{\text{GJ}})}{(3600\ \text{s/hr})(\frac{668}{326}\ \text{seed rate, kg/s})}$$

$(1 - Y_e)$ is, therefore, the required electrostatic precipitator efficiency.

Subappendix AA 9.1.8

POWER COMPARISON BETWEEN A PRECIPITATOR-LEACH SYSTEM AND A CYCLONE-PRECIPITATOR SYSTEM

Precipitator-Leach System

First, the kilograms of water needed to dissolve the seed [rate = 32.71 kg/s K_2SO_4 (71.96 lb/s), ash rate = 3.84 kg/s (8.45 lb/s), Y_c = 0.89, temperature = 541°K (514°F)].

Water to dissolve potassium carbonate

$$(0.89)\ \left(\frac{32.71}{174}\right) \left(\frac{138}{0.567\ \dfrac{\text{kg } K_2SO_4}{\text{kg } H_2O}} \right) = 40.7\ \frac{\text{kg } H_2O}{\text{s}}$$

Water to dissolve potassium sulfate

$$(0.11) \left(\frac{32.71}{0.24\ \dfrac{\text{kg } K_2SO_4}{\text{kg } H_2O}} \right) = 15.00\ \frac{\text{kg } H_2O}{\text{s}}$$

The water to dissolve the potassium carbonate is the controlling factor. The temperature of the seed-water solution is

$$(33.1)\ (0.11)\ \left(\frac{32.71}{174}\right)\ (541 - T) + 29.9\ (0.89)\ \left(\frac{32.71}{174}\right)\ (541 - T)$$

$$+ 13.36\ \left(\frac{3.84}{71.65}\right)\ (541 - T) = (T - 373)\ 40.7$$

or

$$T = 396°K$$

The steam produced is

$$\left(\frac{40.7}{540}\right)\ (396 - 373) = 1.73 \text{ kg/s}$$

The steam needed is then

$$(40.7 - 1.73)\ \frac{540}{568} = 37.05\ \text{kg/s}$$

With the steam enthalpy at 2378 kJ/kg (1022.4 Btu/lb), and with the base case coal having a heating value of 25036 kJ/kg (10766 Btu/lb), the extra kilograms of coal needed is

$$\left(\frac{(37.05)\ (568)}{5981}\right) = 3.518\ \frac{\text{kg coal}}{\text{s}}$$

The power for the pumps is determined on the basis of 40.7 kg (89.54 lb) of water per second, at 60% efficiency, with a head of 1773 kPa (17.5 atm). The power requirement is then about 990 kW.

Cyclone-Precipitator System

For a cyclone, the only power loss is due to the pressure drop (11.1 kPa) (0.11 atm), for a flow of 5.75 m^3 (203 ft^3) [(1.469 MPa (14.5 atm) and 541°K (514°F)], and a compressor efficiency of 50%. This power loss is about 190 kW.

Appendix A 9.2

OPEN-CYCLE MHD PRIMARY HEAT EXCHANGER SIZE AND WEIGHT ESTIMATES

A 9.2.1 Introduction

The heat recovery exchangers for the open-cycle MHD systems are necessarily intimately connected with the plant physical structure and present interesting challenges to the designer. In general, firing to the highest possible temperature and recovering heat from the MHD duct exhaust represent the most efficient cycle options.

In this study, duct exhaust temperatures of 2300°K (3680°F) are typical, with perhaps 10 to 20% slag carry-over. Potassium or cesium seed material is also present. Even though much of the slag and all of the seed is in the vapor state at these conditions, it must be recognized that vaporous slag and seed will diffuse through any refractory towards cooler temperatures and may cause bursting by solidification, slump by viscous phase formation, and erosion by formation of volatile or liquid species.

Immediately, the designer turns to review existing technology in industrial furnaces and high-temperature heat exchange. Smelting, glass and refractory manufacture, and blast furnace stoves represent the analogous industrial processes and equipment. Conditions comparable to the worst slag conditions are found in smelting, but the existing solutions generally involve extensive down times and reconstruction. Comparable temperatures are found in glass and refractory manufacture with acceptable equipment life, but this is a clean fuel technology relying on high-cost fuels. Although blast furnace stoves approximate conditions in MHD heat recovery, typical operations are at lower temperatures, 1800°K (2780°F). In addition, the steel industry sustains considerable annual loss in sensible heat of blast furnace fuel gas, in pumping costs, and in

maintaining equipment to wet-alkaline scrub the blast furnace and coke-oven gas before combustion takes place in the stove. Wet scrubbing eliminates sulfur, alkali, and dust to a concentration (Reference 9.22) of between 0.46 and 0.046 g/m^3 (0.2 and 0.02 gr/ft^3). With a dust loading of 0.46 g/m^3 (0.2 gr/ft^3), approximately 11.43 cm (4.5 in) square flues are required (Reference 9.23); and with 0.046 g/m^3 (0.02 gr/ft^3), 5.1 cm (2.0 in) square flues may be used. The cost per unit volume of a regenerator is much less for the smaller flue sizes.

The argument may be made that the slag which condenses and adheres in the checkers at lower temperature levels may be purged by proper thermal cycling, but this requires very high-temperature duty brick throughout the stove, with resultant high initial cost, and provides no guarantee that the refractory insulating brick will withstand alkali erosion and burst problems. Clearly, the heat recovery exchangers require innovative design. The recovery heat exchanger, secondary combustion system, and seed quench systems might be combined with the diffuser and exhaust duct in a synergistic configuration, affording a degree of protection to the basic structure and achieving maximum efficiency. For example, no known diffuser material can survive in equilibrium with the high-temperature, high-velocity, alkali-slag environment. By use of low-temperature combustion air coolant and a ceramic-coated flange tube wall, a surface is provided that will achieve erosion-deposition equilibrium with the flow. The thickness of the ceramic-slag coating is determined by thermal impedance ratio and slag solidification temperature.

A 9.2.1.1 Recovery Heat Exchanger

Combustion products proceed from the diffuser to the exhaust ducting. In this duct secondary combustion air and seed quench air may be injected. The seed and slag undergo phase change from vapor to solid before the seed is extracted. Considering the problems associated with slag and alkali, the heat exchanger structure must be protected from the corrosive exhaust products. A radiant heat exchanger is used to exchange heat through a slag and alkali free protective boundary layer formed by

secondary combustion air injection or combustion product recirculation. Air is preheated in the radiant exchanger to 1590°K (2402°F), using ceramic tubes above 1370°K (2006°F) gas temperature and various metals below. This choice of temperature is somewhat arbitrary - a high metal temperature being chosen as the less uncertain art and pushed to the material limit.

Flanged tube walls are used as in the diffuser. These have several advantages. Tube supports and pressure tubes are protected from corrosion by a low-stress, sacrificial material that may be patched by welding. Slots for blowing are formed naturally by exchanger fabrication, and the tube wall blowing system protects the exhaust duct wall structure.

Several material candidates are available for the ceramic tubes, depending on the environmental conditions. Silicon carbide,* silicon carbide-coated graphite,* and alumina are considered. It is well known (Reference 9.24) that silicon carbide and alkali metals form high vapor pressure compounds at elevated temperatures. The selection of silicon carbide for exposed surface, high-temperature tubes assumes that the blown boundary protection system is very effective.

Note that for Base Case 3 the preheating of fuel gas containing hydrogen to 1700°K (2600°F) is marginal (Reference 9.27) for hydrogen attack on silicon carbide. At 2028°K (3190°F) silicon carbide will surely fail in service. Base Case 3 Points 4 and 5 are computed for cycle information only and are not presented as a viable design.

A 9.2.1.2 Separately Fired Air Heater (SFAH)

Some of the open-cycle MHD configurations require a separately fired air heater to deliver combustion air in excess of 1590°K (2402°F). In this application, individually fired regenerators of the blast furnace stove type are recommended. The fuel is low- to medium-Btu coal gas with about 0.57 g/m^3 (0.25 gr/ft^3) dust loading. Despite previous data indicating that 11.4 cm (4.5 in) square flues are required with low-Btu gas, these stoves are designed on the basis of 5.1 cm (2.0 in) square flues. A slightly higher than normal replacement rate in the O&M charges is assumed.

*Thermal cycling increases the oxidation rate of SiC, Trinks (Reference 9.34).

In addition, straight flues without lateral passages are used to minimize dust accumulation. At various elevations the checkers are supported by arches, both to reduce checker bottom stress and to allow redistribution of flow into blocked flues.

The separately fired combustion air preheater (GAP) is a recuperative, muffle type. Costs of this device were indicated to be exceptionally high,so this design may not be near optimum.

A 9.2.1.3 Costing

Costing performed at the conceptual design level consisted of determining the quantity of materials in a basic device and structure without considering codes or service; estimating labor, material, and manufacturing costs; and breaking them down into per unit amounts.

A 9.2.2 Summary

The open-cycle MHD primary heat exchangers represent a considerable extrapolation of current technology. The following are the main areas of concern:

- Effectiveness of the blown boundary layer protection system
- Suitability of silicon carbide exposed surface tubes for the required duty
- Upper metal temperature limit or ceramic tube crossover point
- Stove flue size and material for the specific combustion gas impurities encountered.

A 9.2.3 Recommendations

- The blown boundary layer protection system effectiveness should be determined by a thorough experimental program.
- A suitable ceramic tube should be proof tested.

A 9.2.4 Diffuser Analysis and Design

Recovery of the MHD duct dynamic pressure is accomplished in a three-dimensional plane wall diffuser with 7.5 degree half angles. It is

perhaps optimistic to assume high-efficiency recovery with slag-covered rough walls with 7.5 degree half-angle divergence;but cooling should help, and this is not fundamental to the cycle performance. Figure A 9.2.1 shows the ceramic-lined flange tube wall cross section, and Figure A 9.2.2 shows the diffuser circuit schematic. The diffuser design assumes that gunned, chrome-bonded alumina will withstand the alkali vapor, velocity, and slag at 1778°K (2740°F) for extended periods. From a thermal impedance ratio, the refractory thickness is found to be about 1.27 cm (0.5 in).

A 9.2.4.1 Diffuser Dimensions

The following are the reference diffuser design conditions:

$$V_o = 30.5 \text{ m/s } (100 \text{ ft/s})$$

$$\rho_o = 0.183 \text{ kg/m}^3 \ (0.01142 \text{ lb/ft}^3)$$

$$\dot{m} = 1426 \text{ kg/s } (3143.7 \text{ lb/s})$$

These conditions are abstracted from Base Case 2, Point 1,MHD generator design program computer output dated 2/25/75. The program assumes adiabatic diffuser walls. The diffuser outlet dimension is taken from the conservation of mass equation expressed as:

$$S_o = \left[\frac{\dot{m}}{\rho_o V_o}\right]^{1/2} = \left[\frac{1426}{(0.183)(30.48)}\right]^{1/2} = 15.9 \text{ m} \qquad \text{(A 9.2.1)}$$

for a square cross-sectional duct. Substitution of the assumed density mass flow rate and flow velocity into Equation A 9.2.1 yields a value of S_o of 15.9 m (52 ft). The axial diffuser length is given for 7.5 degree expansion half-angles as Equation A 9.2.2:

$$L = \frac{S_o - S_i}{2 \tan 7.5°} \qquad \text{(A 9.2.2)}$$

S_i is taken from the above mentioned computer printout as 3.9 m (12.8 ft), yielding a diffuser length, L, of 44.8 m (147 ft). These dimensions determine the irradiated surface of the diffuser.

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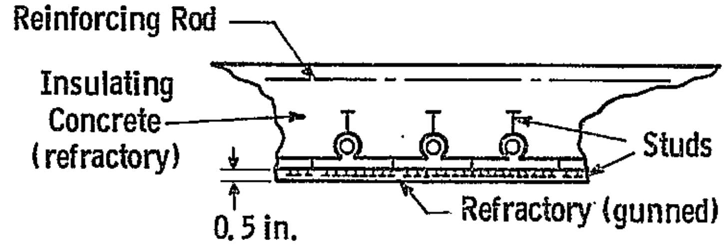

Fig. A 9.2.1—Ceramic lined flange tube wall of the diffuser

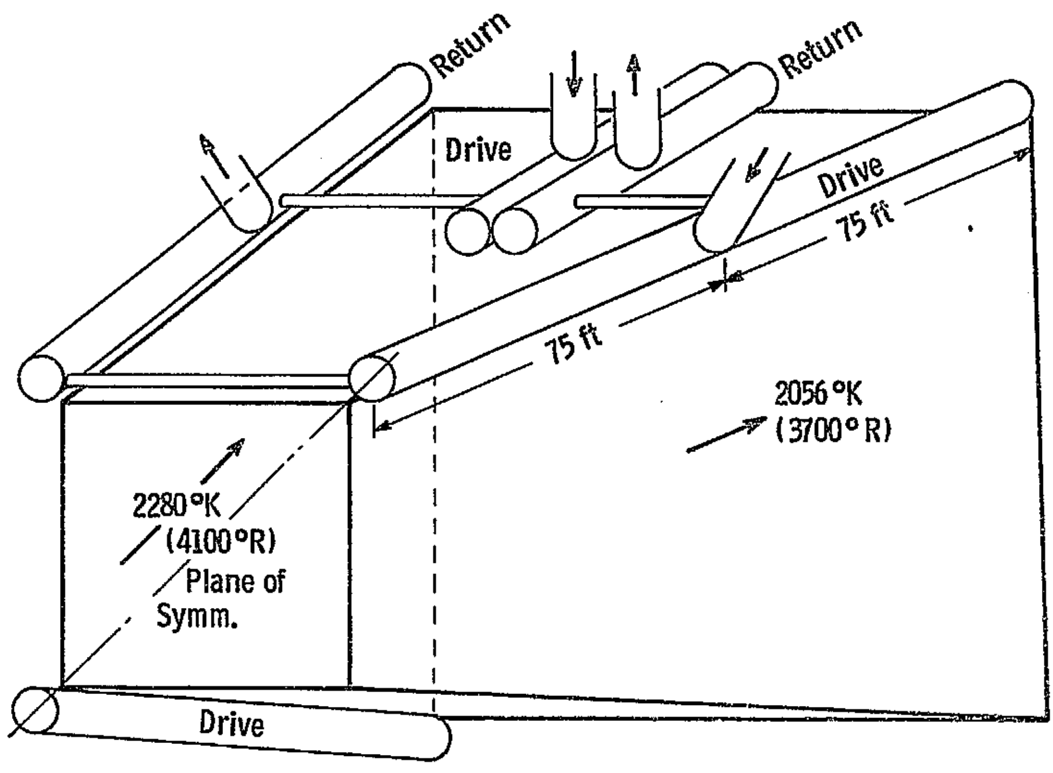

Fig. A 9.2.2—Diffuser tube and header circuitry schematic

9.2.4.2 Diffuser Thermal Analysis

Heat is delivered to the diffuser wall by both radiation and convection. The radiation and convection heat transfer coefficients h_r and h_c are given by Equations A 9.2.3 and A 9.2.4.

$$h_r = \frac{\sigma(\varepsilon_G T_G^4 - \alpha_{G1} T_1^4)}{\Delta T_m} \qquad \text{(A 9.2.3)}$$

$$h_c = 0.037 \frac{k_G}{L} Re_L^{0.8} Pr^{0.33} \qquad \text{(A 9.2.4)}$$

Equation A 9.2.3 is recognized as the radiant exchange coefficient between a radiating gas at T_G and black walls at T_1. Equation A 9.2.4 is the turbulent flat plate convection coefficient. The assumption of black surfaces in the diffuser is conservative; and since the equilibrium roughness and material are not known, the assumption of unity emissivity is justified. The assumption of the turbulent flat plate boundary layer with an adverse pressure gradient is also conservative and is justified, since surface roughness can only be estimated as less than the laminar sublayer thickness.

A significant problem in radiative heat exchange is the determination of α_{G1}. If $T_G = T_1$, then $\alpha_G = \varepsilon_G$. But this is not the case. McAdams (Reference 9.25) recommends absorptivity change of the gas due to molecular and density effects be accounted for by determining α_{G1} as $\varepsilon_G (T_G/T_1)^{0.65}$, where ε_G is the emmissivity for a temperature T_1 and effective partial pressure-length factor of $PL(T_1/T_G)$. For the temperatures shown in Figure A 9.2.2 the assumption that $\alpha_G = \varepsilon_G$ is conservative. The value ε_G is formulated as the sum of emissivities of the emitting species, carbon dioxide and water vapor in these cases, with appropriate correction factors for overlapping emission wavelength bands. Emission from the vapors of potassium sulfate, particulates, carbon monoxide, slag vapors, and the like are considered negligible. This nonconservative assumption tends to offset the conservative black wall assumption. The gas emissivity

is given as:

$$\varepsilon_g = \varepsilon_{CO_2} + \varepsilon_{H_2O} - \Delta\varepsilon_{CO_2,H_2O} \qquad \text{(A 9.2.5)}$$

For Illinois No. 6 coal fired in the as received condition with 0.95 stoichiometric air, the partial pressures of the carbon dioxide and water vapor are calculated to be 15.81 and 10.34 kPa (0.156 and 0.102 atm), respectively.

The characteristic radiating length in the diffuser is given by Equation A 9.2.6:

$$L_o = \frac{4V}{A_s} = 33 \text{ ft} \qquad \text{(A 9.2.6)}$$

Although correction factors, L/L_o, to this basic length for the diffuser geometry are not available, a good estimate from Table 4-3 of McAdams (Reference 9.25) is given by 0.85 and 0.77 for the water vapor and carbon dioxide, respectively. The radiation opacity terms are given as Equation A 9.2.7 and A 9.2.8.

$$(PL)_{CO_2} = 0.156\ (33)(0.77) = 3.96 \text{ ft-atm} \qquad \text{(A 9.2.7)}$$

$$(PL)_{H_2O} = 0.102\ (33)(0.85) = 2.80 \text{ ft-atm} \qquad \text{(A 9.2.8)}$$

Sufficient data exist to enter Figures 4-13, 4-15, and 4-17 of Reference 9.25 to determine ε_g from Equation A 9.2.5.

$$\varepsilon_{G_{4100°R}} = 0.115 + 0.158 - 0.058 = 0.215 \qquad \text{(A 9.2.9)}$$

By similar reasoning

$$\alpha_{G1} = 0.160 + 0.207 - 0.058 = 0.309 \qquad \text{(A 9.2.10)}$$

In order to iterate the design it is useful to cast the radiation heat exchange into an effective coefficient form, as in Equation A 9.2.3:

$$h_r = \frac{0.1714\ (0.213\ (41)^4 - 0.309\ (32)^4)}{(4100 - 3200)} = 54 \text{ Btu/hr-ft}^2\text{-°F} \qquad \text{(A 9.2.11)}$$

The values assumed for the calculation of the convection coefficient were:

$\rho = 0.165\ kg/m^3\ (0.0103\ lb/ft^3)$

$V = 330\ m/s\ (1083\ ft/s)$

$L/2 = 24\ m\ (78.74\ ft)$

$\mu = 5.787\ Ns/m^2\ (0.14\ lb/ft\text{-}hr)$

$k = 0.1125\ W/M\text{-}°K (0.065\ Btu/hr\text{-}ft\text{-}°F)$

$Re_{L/2} = 23 \times 10^6;\ Pr_{0.33} \approx 1.0$

A length dimension equal to 0.5 times the diffuser length was chosen as conservative. Substitution of these properties into Equation A 9.2.4 gives the estimated convection coefficient, h_c as 130.6 $W/m^2\text{-}°K$ (23 $Btu/hr\text{-}ft^2\text{-}°F$). The gas-to-wall total coefficient is estimated using Equation A 9.2.12.

$$U_{c+r} = h_r + h_c = 77\ Btu/hr\text{-}ft^2\text{-}°F \qquad \text{(A 9.2.12)}$$

The effective conductance from the refractory surface to the cooling air is determined iteratively. The cooling air leaves the compressor at 0.6586 MPa (6.5 atm) and 514°K (465°F). From the known surface temperature and irradiated diffuser area, the temperature rise of the air may be found to be approximately 245°K (441°F) and the diffuser outlet combustion products temperature, 1970°K (3086°F).

After several iterations it is found that 5.08 cm (2.0 in) Schedule 40 pipes with flanges 0.635 cm (0.25 in) thick by 7.66 cm (3.0 in) wide as in Figure A 9.2.1, arranged to have 3600 parallel circuits, represent a satisfactory design compromise. Calculations from the final iteration are abstracted to indicate the procedure.

The wall conductance, U_t, to keep the refractory surface at T_1 [1778°K (2740°F)] is represented as Equation A 9.2.13, where $\bar{T}_a$ is the average air temperature in the diffuser wall cooling passages:

$$U_{c+r} \quad (T_G - T_1) = U_t \; (T_1 - \bar{T}_a) \qquad \text{(A 9.2.13)}$$

$$U_t = \frac{77(900)}{(3200-1100)} = 33.0 \text{ Btu/hr-ft}^2\text{-°F} \qquad \text{(A 9.2.14)}$$

U_t', the heat transfer coefficient per foot of tube based on a unit area of refractory surface is conservatively approximated by Equation A 9.2.15),

$$U_t'A_f' = \cfrac{1}{\cfrac{1.0}{\eta_t \; \pi d_i \; h_i} + \cfrac{\ln(d_o/d_i)}{2\pi \; k} + \cfrac{t_{(Cr\;O_2-Al_2O_3)}}{\eta_f \; A_f' \; k_{(Cr\;O_2-Al_2O_3)}}} \qquad \text{(A 9.2.15)}$$

and since four tubes per square foot of irradiated surface are assumed, A_f' is 0.0762 m^2/m (0.25 ft^2/ft). U_t' is equal to 0.25 U_t or 14.28 W/m-°K (8.25 Btu/hr-ft_{tube} -°F).

For the stated tube-side air conditions of 0.6586 MPa (6.5 atm) and 514°K (465°F), the perfect gas law gives the air density.

$$\rho = \frac{p}{RT} = \frac{6.5 \; (14.7)(144)}{53.3 \; (925)} = 279 \; \frac{lb}{ft^2} \qquad \text{(A 9.2.16)}$$

Assuming an air velocity of 38.1 m/s (125 ft/s) in the 5.08 cm (2 in) diameter tubes, the Reynolds number of the flow is given by Equation A 9.2.17:

$$Re_d = \frac{\rho V d}{\mu} - \frac{0.279 \; (125)(2/12)}{2.0 \times 10^{-5}} = 2.91 \times 10^5 \qquad \text{(A 9.2.17)}$$

For an air conductivity, k, of 0.0465 W/m-°K (0.0269 Btu/hr-ft-°F), and assuming the Prandtl number raised to the one-third power is approximately one, the heat exchange coefficient in the tube is given by Equation A 9.2.18.

$$h_i = 0.023 \frac{k}{d} \mathrm{Re}^{0.8} \mathrm{Pr}^{1/3} = 0.023 \frac{0.0269}{(2/12)} (2.91)^{0.8}\, 10^4 \quad \text{(A 9.2.18)}$$

$$= 87.1 \text{ Btu/hr-ft}^2\text{-°F}$$

The first two terms of Equation A 9.2.15 are evaluated assuming $\eta_t = 0.8$.

$$\frac{1}{\eta_t \pi d_i h_i} = \frac{12}{0.8(\pi)(2)87.1} = 0.0274 \ (\text{Btu/hr-ft-°F})^{-1} \quad \text{(A 9.2.19)}$$

$$\frac{\ln(d_o/d_i)}{2\pi k_t} = \frac{\ln(\frac{2.375}{2})}{2\pi(14)} = 0.00195 \ (\text{Btu/hr-ft-°F})^{-1} \quad \text{(A 9.2.20)}$$

The refractory thickness is found from Equation A 9.2.15.

$$\left.\frac{t}{\eta_f A'_f k}\right|_{CrO_2-Al_2O_3} = \frac{1}{8.25} - 0.0274 - 0.00195 = 0.092 \quad \text{(A 9.2.21)}$$

Assuming η_f is 0.85 and substituting A'_f as 0.0762 m^2/m (0.25 ft^2/ft) and k for $CrO_2 - Al_2O_3$ from Table A 9.2.1 as 2.942 W/m-°K(1.7 Btu/hr-ft^2-°F) yields a refractory thickness, t, of 1.01 cm (0.40 in). A thickness of 1.27 cm (0.5 in) was used to allow for erosion-deposition stabilization. Stainless steel was used behind the refractory to provide some corrosion resistance in case of refractory spalling.

Table A 9.2.1 Refractory Properties

Temperature, °F From	To	Refractory	ρ lb/ft^3	C_s Btu/lb-°K	k Btu/hr-ft-°K	\$/lbm*
	2300	X.H.D. fireclay	145	0.264	0.83	0.164
2300	2700	Mullite	155	0.252	1.0	0.407
2700	3100	High alumina	185	0.278	1.58	0.463
3100	3500	Chrome bonded alumina or calcia stabilized zirconia	185	0.278	1.7	0.482
3500	4100	Yttria stabilized zirconia	322	0.17	1.0	{3.30 7.00

The MHD loop computer design program assumes adiabatic diffuser walls, but from this point in the thermal analysis onward the thermodynamic state of the various working fluids may deviate from the computer results as actual heat flows are accounted for. Heat flow through the diffuser wall into the primary combustion air is computed by an energy balance between the air and products of combustion. The overall (UA)' is computed using Equation A 9.2.22

$$(U_t{}'A_f') = \frac{1}{\dfrac{1}{\eta_t \pi h_i d_i} + \dfrac{\ln(d_o/d_i)}{2\pi k_t} + \dfrac{U_{r+c} + k/t}{\eta_f A_f{}' \; U_{r+c} \; k/t}} \qquad \text{(A 9.2.22)}$$

where k and t both refer to the thermal conductivity and thickness of the gunned refractory. Substitution of properties already given in this section results in an overall (UA)' of 35.36 W/m^2-°K (6.22 Btu/hr-ft-°F). Although the solution is iterative, only the final interation is given here. The heat transferred through the wall for an assumed duct length of 44.8 m (147 ft) and a mean duct dimension of 9.75 m (32 ft) is:

$$\dot{q} = (UA)' \; L \; \Delta T_m$$

$$\dot{q} = (6.22)\ [(\frac{1}{0.25})\ (4)\ (32)\ (144)]\ (3800 - 1100) \qquad (A\ 9.2.23)$$

$$= 1.264 \times 10^9 \text{ Btu/hr}$$

In terms of the relative enthalapy of the 2502 MHD Thermochemical Properties of Combustion Gases Computer Program run for as received Illinois No. 6 coal dated 2/17/75, the diffuser outlet products have a relative enthalpy of:

$$i = -0.565\ [MJ/kg] - \frac{1.27 \times 10^9\ [\ \text{Btu/hr}]\ 1054[J/Btu]}{1426[kg/s]3600[s/hr]} \qquad A\ 9.2.24)$$

$$i = -0.826\ [MJ/kg]$$

Table A 9.2.2 Temperature-Enthalpy for Combustion Products

T, °K	i,MJ/kg
1900	-0.940
1970	-0.826
2000	-0.783

From the interpolation in Table A 9.2.2, the outlet temperature of the diffuser air is 1970°K (3086°F) This is sufficiently close to the assumed outlet of 3680°R in Figure A 9.2.2 to terminate the iteration.

To complete the design, 3600 parallel tubes are required, as in Equation A 9.2.25.

$$n_t = \frac{1248\ [kg/s]2.2[lb/kg]}{125\ [ft/s]0.279\ [lb/ft^3]0.0218\ [ft^2/tube]} \qquad \text{(A 9.2.25)}$$

$$\approx 3600$$

For 3600 parallel circuits, the average tube length is less than 7.62 m (25 ft). as seen in Figure A 9.2.1.

Pressure drop, neglecting headers, in the diffuser cooler is given as:

$$\Delta p = \left\{ f\frac{\ell}{d} + C_L \right\} \rho \frac{V^2}{2g_c} = \left\{ 0.025\ \frac{25(12)}{2.0} + 1.0 \right\} \frac{(0.279)(125)^2}{(2)(32.174)(144)}$$

$$\Delta p = 2.23 \text{ psi} \qquad \text{(A 9.2.26)}$$

A 9.2.4.3 Diffuser Scaling

The MHD duct outlet temperature for all cases, is nearly constant, as is the compressor outlet temperature. The products-to-combustion air mass ratio is also a second-order effect on mean temperature difference. Given these constraints, scaling of the diffuser dimensions and material requirements is accurately and simply represented as

$$A_{H.X.} = 2\ (S_i + 1.46\sqrt{\dot{m}_a}\ )(2.82)(1.46)\sqrt{\dot{m}_a} \qquad \text{(A 9.2.27)}$$

The numerical value 1.46 has the units $ft/(kg/s)^{1/2}$ and is obtained from Equation A 9.2.28 and the combustion airflow rate is:

$$\frac{52\ [ft]}{\sqrt{1256\ kg/s}} = 1.46 \qquad \text{(A 9.2.28)}$$

The numerical value 2.82 is a dimensionless proportionality constant.

It is calculated by assuming the diffuser length is proportional to the outlet dimension.

$$\frac{L}{S_o} = \frac{147}{52} = 2.82 \qquad \text{(A 9.2.29)}$$

Note that for the lower MHD duct flow rates the diffuser tube diameter is reduced to maintain pressure drop simular to the base case. The incremental change in tube-side thermal impedance is negligible.

A 9.2.4.4 Diffuser Material Content

The materials per foot^2 of diffuser surface summed from the previous dimensions are shown in Table A 9.2.3.

Table A 9.2.3 - Diffuser Material Content

Ruby refractory 1/2 in

$$\frac{1}{24}\, 1\ [ft^3]\ (197)\ [\frac{lb}{ft^3}] = 8.2\ lb/ft^2$$

Flange

$$\frac{1}{48}\,(1)\ (1)(1728)\ (.289) = 10.4\ lb/ft^2$$

Tube

$$4\ (1)\ (1.075)\ (.289)\ (12) = 14.9\ lb/ft^2$$

Σ Flange and tube 25.3 lb/ft^2

Headers

10% of flange-tube = 2.53 lb/ft^2

10% for headers is a factor obtained from a cost analysis of this diffuser (Figure A 9.2.2) design.

Diffuser Structural Material

The diffuser slab is presumed to be of reinforced concrete 0.4572 m (1.5 ft) thick and trapezoidal in plan. For the computed design 206 m^3 (270 yd^3) of concrete are required. To this are added 6.88 m^3 (9 yd^3) of concrete for an expansion joint foundation, making the total 213.3 m^3 (279 yd^3). Materials given in Table A 9.2.4 are for a refractory lined steel shell with external framing, even though refractory concrete may represent a better choice.

Table A:9.2.4 Diffuser Structural Materials

Structural Steel	20,000 lb
Plate Shell, 1/2 in	408,000 lb
Refractory, 12 in. thick for 190°F cold face	860,000 lb
Anchor Bolts	$3.00/$ft^2$ H.X. surface
(Mat'l and labor included in insulation cost)	
Reinforced Concrete	279 yd^3

For other parametric points each of the above material quantities is assumed proportional to the area of heat exchange (H.X.) surface.

A 9.2.4.5 Diffuser Operating and Maintenance Charges

Although no data exist for gunned chrome-bonded alumina refractory surfaces, it is arbitrarily assumed that 20% of the diffuser surface must be sand blasted and regunned each year. Costs are given in Table A 9.2.5.

Table A.9.2.5 Diffuser Maintenance Costs

Operation	Cost (Mat'l & labor), $\$/ft^2$	Direct Charge, $\$/(ft^2 H.X.)(yr)$
Sand Blasting	1.50	0.30
Gunning	3.50	0.70
Welding, (1/2 in. thick mat'l)	3.30	

A 9.2.5 Recuperator: Air Temperature > 1367°K (2000°F)

The recuperator is immediately downstream of the diffuser in the products stream. At the time the recuperator estimate was prepared some uncertainty existed about whether the secondary combustion air would be injected immediately downstream of the diffuser, injected downstream of the recuperator, or injected as a protective boundary layer in the recuperator. It was decided that the design would be based on the MHD duct computer program temperatures and later scaled to account for the nonadiabatic diffuser and secondary combustion air injection.

The temperature distance chart (Figure A 9.2.3) shows that the recuperator must be a counterflow device. Exhaust products enter the exhaust duct at 2056°K (3240°F) and air leaves the ceramic tube section at 1589°K (2400°F). From the air inlet condition of 0.6282 MPa (6.2 atm) and 1367°K (2460°R), the heat exchange is given as

$$\dot{q} = \dot{m}\,(i_o - i_i) \tag{A 9.2.30}$$

As stated previously, silicon carbide in the presence of alkali and sulfur vapor and air forms volatile species and, therefore, must be protected. Upstream from the recuperator is an annular separator which shunts liquid slag from the top and side walls to the floor of the exhaust duct. The tubes are protected by a blown boundary layer of alkali-free air or recirculated exhaust products. Figure 9.2.4 shows a cross section of an extruded ceramic tube. Wall thickness is computed using

Curve 681142-A

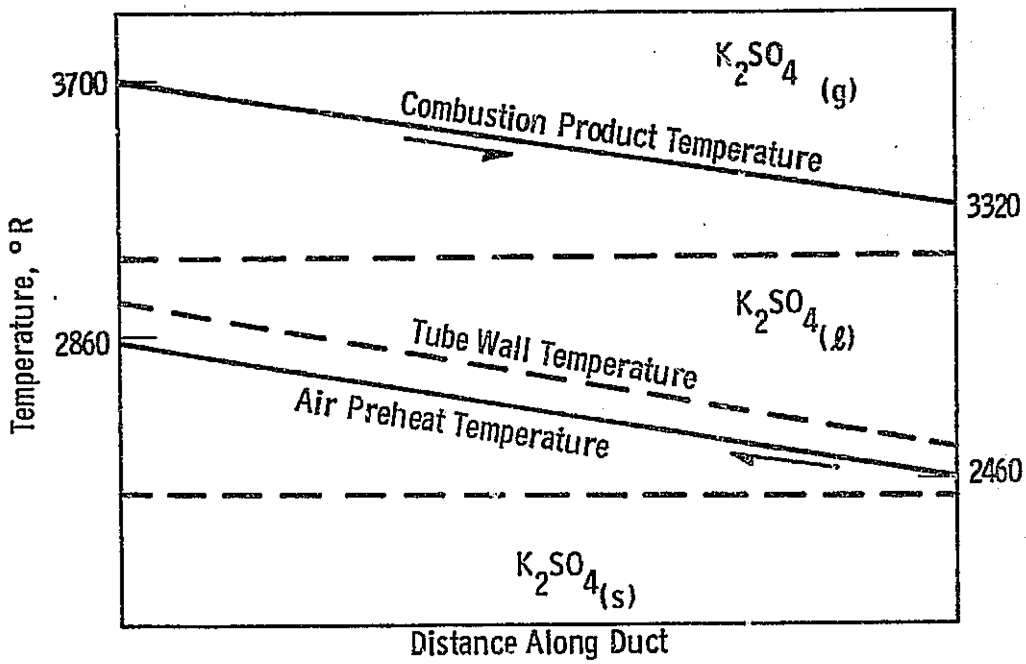

Fig. A 9.2.3—Temperature distance chart for the ceramic tube section of the recuperative heat exchanger

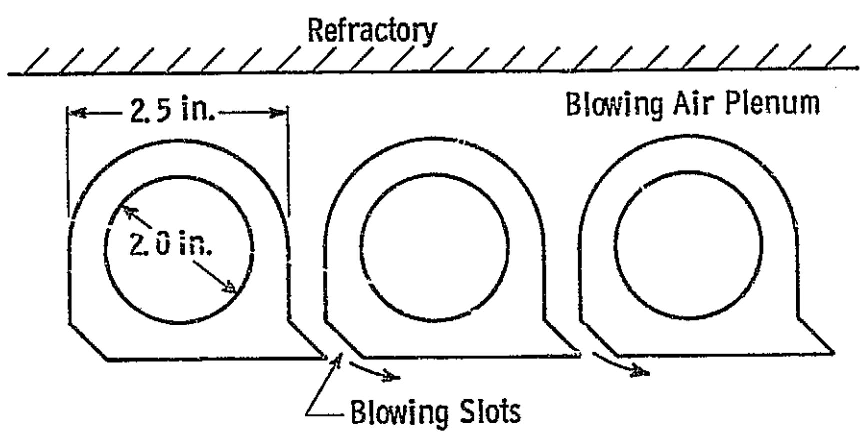

Fig. A 9.2.4—Cross section of extruded ceramic tubes

stress and creep for dense graphite, presuming silicon carbide surface treatment of graphite is an acceptable alternative material process. In any case, graphite tensile strength lies well below that of silicon carbide (KT), SiC-Si (Durhy) and $SiC-Si_3N_4$ (Refax) (Reference 9.27). The required wall thickness for thermal design purposes is calculated assuming thin wall stress distribution according to the Equation A 9.2.31:

$$t = \frac{Pd}{2\sigma} \qquad \text{(A 9.2.31)}$$

For 5.08 cm (2 in) id graphite (thinner walls for silicon carbide) tubes and an allowable stress of 3.447 MPa (500 psi), a wall thickness of 4.7 mm (0.185 in) is required to withstand a design pressure of 0.638 MPa (92.5 psi). A 6.35 mm (0.25 in) wall thickness seems reasonable, since the surface most likely to erode is very thick.

The overall thermal conductance per foot of tube is given as

$$(UA)' = \frac{1.0}{\dfrac{1.0}{\eta_t \pi d_i h_i} + \dfrac{\ln(d_o/d_i)}{2\pi\ k_t} + \dfrac{1}{A'\ U_{c+r}}} \qquad \text{(A 9.2.32)}$$

For an average air velocity of 38.1 m/s (125 ft/s), the Reynolds number based on average velocity and properties is 5.4×10^4.

The Colburn correlation, Equation A 9.2.33 (Reference 9.28) yields a tube-side heat transfer coefficient of 232.8 W/m^2-°K (41 Btu/hr-ft^2-°F).

$$h_i = 0.023 \frac{k_{air}}{d_i} Re^{0.8}\ Pr^{0.33} = 41.0 \text{ Btu/hr-ft}^2\text{-°F} \qquad \text{(A 9.2.33)}$$

Assuming the tube wall surface is 80% thermally efficient, the dimensions of pipe in Figure A 9.2.4, mean properties, and U_{c+r} = 210 W/m^2°K(37 Btu/hr-ft^2°F) gives (UA)' as 35.6 W/m^2-°K (6.28 Btu/hr-ft^2-°F).

The determination of U_{c+r} as 210 W/m^2°K(37 Btu/hr-ft^2-°F) deserves some discussion. Figure A 9.2.4 indicates that the convection component of the combined radiation-convection coefficient is determined by blown boundary

layer theory and local temperature difference to the transpiring stream. In effect, part of the heat delivered to the tube by radiation is lost by convection to the cooler transpiring stream. Thus it is necessary to know the required blowing rate.

Almost perfect protection of the surface is required. In terms of the diffusion mass transport of the vaporous alkali-sulfur species, good protection requires that sufficient blowing be established so that the diffusion coefficient is small. Unfortunately, existing diffusion transpiration theories, Hartnet and Eckert (Reference A 9.29) for laminar flat plate, and Rubesin and Rubesin and Pappas for turbulent flat plate as reported by (Reference A 9.30), Kays, accounts for only molecular diffusion. The effects of laminarization by acute angle injection, convective diffusion, and gravitational separation can only be estimated at this time. The blowing rate is determined by requiring the molecular diffusion coefficient to be reduced by 50% according to Rubesin's approximation.

From Figure 15-3 of Reference 9.30 for $h_D/h_{Do} = 0.5$, similar molecular weights, and turbulent conditions

$$\frac{\dot{m}''}{0.5\ h_{D_o}} = \frac{\dot{m}''}{h_D} = 2.3 \qquad \text{(A 9.2.34)}$$

It should be noted that Equation A 9.2.34 is dimensionless, $\dot{m}''$ having the dimensions of ft^3 gas/(ft^2 surface)(s) for compatibility with Reference 9.30. From the Chilton-Colburn analogy between mass and heat transfer, h_D is estimated from Equation A 9.2.35 for flat plate flow with a Lewis number of 1:

$$h_{D_o} = \frac{0.037\ k}{\rho C_p\ L} (Re_L)^{0.8}\ Pr^{0.33} \qquad \text{(A 9.2.35)}$$

The properties of combustion products are assumed identical to those of air where unknown, and length is conservatively estimated as 15.24 m (50 ft). The parameters used are given below in Table A 9.2.5. Substitution into Equation A 9.2.35 yields h_{D_o} equal to 8.737 cm/s (1032 ft/hr) and substitution into Equation A 9.2.34 gives $\dot{m}''$ equal to 0.33 $ft^3/(ft^2$-H.X.-s).

Table A 9.2.5 Flue Gas Properties

Parameter	Value
L	30 ft
V	100 ft/s
ρ	0.0156 lb^3
μ	0.14 lb/ft-hr
Cp	0.28 Btu/lb-°F
k	0.05 Btu/hr-ft-°F
$Pr^{1/3}$	1.0
Re	1.2×10^6

By analogy the convection coefficient h, locally applicable, is estimated as

$$h = 0.5 h_{D_o} \; \rho \; Cp = 2.25 \text{ Btu/hr-ft}^2\text{°F} \qquad \text{(A 9.2.36)}$$

If acute angle injection is able to return the boundary layer to laminar flow, the blowing rate from (Reference 9.29) for h_D/h_{Do} = 0.5 would be 0.006096 m^3/m^2-s(6.02 ft^3/ft^2-s). Between 0.01829 and 0.06096 m^3/m^2-s (0.06 and 9.2 ft^3/ft^2-s) should be adequate.

The radiation coefficient, by techniques of Equations A 9.26 through A 9.2.12 is found to be 232.8 W/m^2-°K (41 Btu/hr-ft^2-°F). The combined coefficient is formulated as

$$U_{c+r} = 41 - 2.25 \; \frac{(2660-1110)}{850} = 37 \text{ Btu/hr-ft}^2\text{-°F} \qquad \text{(A 9.2.37)}$$

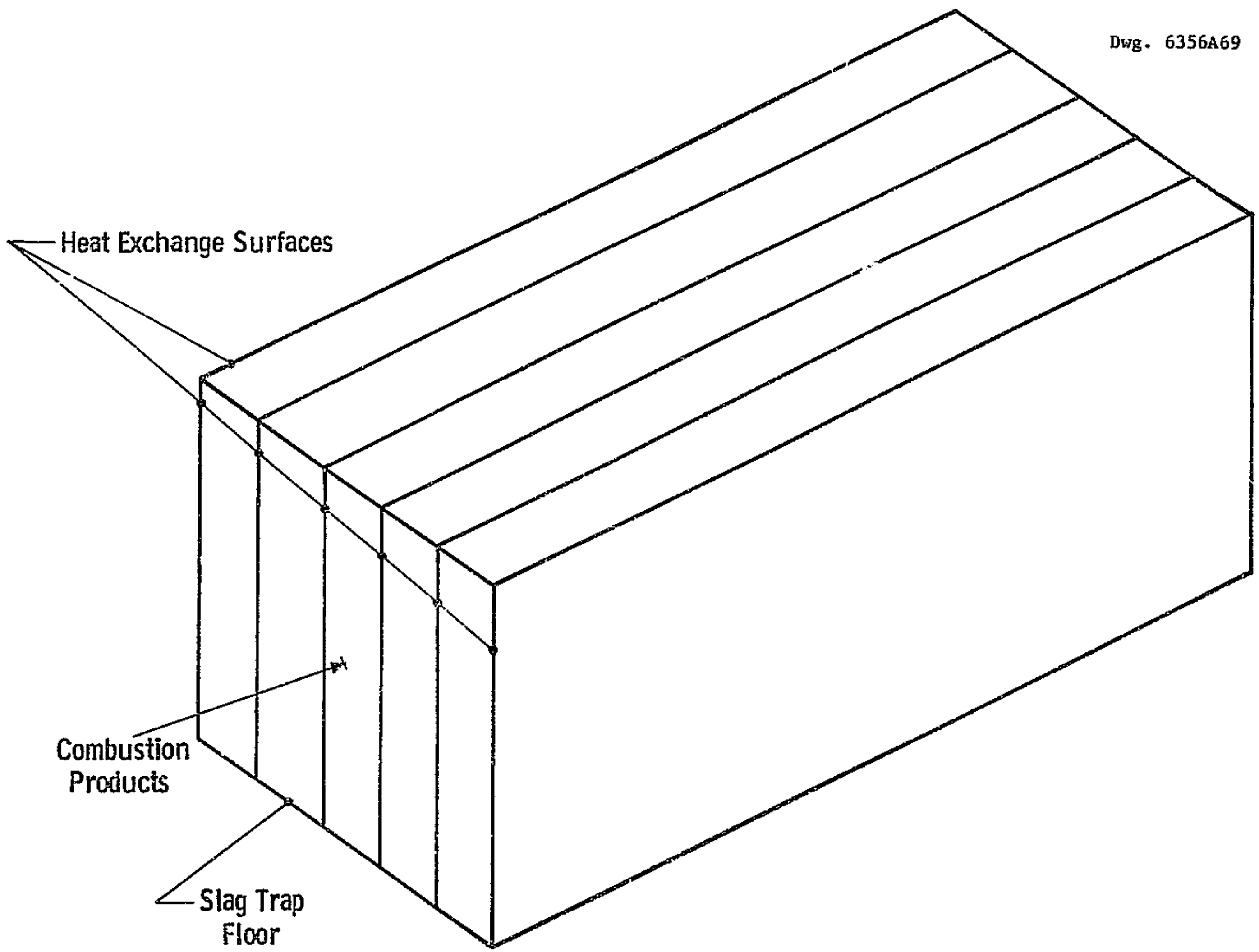

Fig. A 9.2.5—Duct schematic showing splitters

Overall log mean temperature difference is given as

$$\Delta T_{\ell m} = \frac{(3700-2860)-(3320-2460)}{\ell n \frac{(840)}{(860)}} = 850°F \qquad \text{(A 9.2.38)}$$

The enthalpy difference to heat the air from 1367 to 1589°K (2000 to 2400°F) is 0.2687 MJ/kg (115.53 Btu/lb). The heat added in the high-temperature recuperator is given by Equation A 9.2.39:

$$\dot{q} = \dot{m}\ \Delta i = [(1256)\ (3600)\ (\frac{1}{0.4536})]\ (115.53) = 1.151 \times 10^9 \text{ Btu/hr} \qquad \text{(A 9.2.39)}$$

The required length of tube is given by Equation A 9.2.40:

$$L = \frac{q}{(UA)'\Delta T_{\ell m}} = \frac{1.151 \times 10^9}{(6.28)\ 850} = 2.157 \times 10^5 \text{ ft} \qquad \text{(A 9.2.40)}$$

For the geometry shown in Figure A 9.2.4 which has a surface area per unit tube length of 6.35 cm^2/cm (0.208 ft^2/ft), the surface area required is 4175 m^2 (44938 ft^2). Continuity indicates for a velocity of 59.1 m/s (195 ft/s) and 5.08 cm (2 in) id tubes, 8628 parallel circuits are required of 7.62 m (25 ft) length. This represents a convenient circuitry. The tube side h is based on 38.1 m/s (125 ft/s) in Equation A 9.2.33 and is not recalculated.

In order to minimize axial length, four vertical splitters are used in the duct. The correction to mean radiating length is given by McAdams (Reference 9.25, Table 4-2 and 4-3). Figure A 9.2.5 shows the duct schematic in the ceramic tube section. The radiation coefficient previously cited is calculated for the splitters of Figure A 9.2.5. With four vertical splitters, eleven surfaces are available for heat exchange, excluding the slag tap floor.

For a 15.84 m (52 ft) square duct, and assuming 7.62 m (25 ft) lengths of the tube supplied by a header down the center, the silicon carbide-lined duct will be approximately 25 m (82 ft) long.

Pressure drop in the ceramic tube primary heat exchanger is

$$\Delta p = \left(f \frac{L}{d} + C_L\right) \rho \frac{V^2}{2g_c} = 0.75 \text{ psi} \qquad \text{(A 9.2.41)}$$

Joining the extruded ceramic tubes to the header is accomplished by platinum-plated molybdenum metal "V" rings which allow for thermal expansion and simple replacement.

A 9.2.5.1 Ceramic Recuperator Scaling

Several characteristics permit scaling the ceramic recuperator to other conditions. The design and tube circuitry are flexible enough to apply to all cases. In addition, the ceramic recuperator is used above 1367°K (2000°F) in all cases. From Equation A 9.2.40 the heat exchange area is proportional to heat rate and inversely proportional to log mean temperature difference, (UA)' being independent of velocity of the products. If the log mean temperature difference is approximated as the average temperature difference, the required surface area, A_s, is given by the following proportion:

$$A_s \propto \frac{\dot{m}_A \left(T_{a_o} - 2460\right)}{T_{p_i} - T_{a_o} + T_{p_o} - 2460} \qquad \text{(A 9.2.42)}$$

A 9.2.5.2 Ceramic Recuperator Structural Materials

The slab, shell, and structural steel of the ceramic recuperator are similar to the metal recuperator, so they are computed as a unit in Section 9.2.6. Operating and maintenance charges were assumed to be the same per unit area as in the metallic recuperator and also are given in Section 9.2.6

A 9.2.6 Metallic Recuperator

As presented in the introduction, the change from metal to ceramic tubes is arbitrarily set at 1367°K (2000°F). The hot end tube-wall temperature is then about 1422°K (2100°F). Design data for the proposed material, RA 333, at these conditions are tentatively taken from the manufacturers' catalog (Reference 9.31). RA 333 is an alloy composed of 45% nickel; 25% chromium; 18% iron; 3% each of tungsten, cobalt, and molybdenum; 1.25% silicon; and 2.0% maganese. At 1422°K (2100°F) the stress to produce 2.78×10^{-8} %/s (0.0001%/hr) secondary creep is 965 kPa (140 psi) (Reference 9.31). The alloy does not resist sulfur well at high temperature, so a protective boundary layer is used; less protection is required at lower temperatures. In a creep stress analysis, uniform stress distribution may be assumed. The wall thickness is determined using Equation A 9.2.31. At the hot end the pipes are assumed to have a 1.68 cm (0.66 in) wall thickness. The wall thickness of the 5.08 cm (2 in) pipe increases with temperature from 0.554 cm (0.218 in) to 1.68 cm (0.66 in). The greater portion of the exchanger pipe has thin walls so the smaller dimensions are used to estimate the resistance to heat flow.

Analysis proceeds as in the ceramic tube section with new material properties, emissivities, and temperature differences. Figure A 9.2.6 is a counterflow temperature distance chart for the metallic recuperator. Assuming 83°K (150°F) for air-to-tube temperature difference leads to evaluating the average tube thermal conductivity from Reference 9.3.1 as $\bar{k}_{t_{1900°R}}$ equal 19.9 W/m-°K (11.5 Btu/hr-ft-°F). In evaluating the radiation coefficient, the radiation opacity parameters, P_pL in ft-atm, are found to be 1.44 and 2 for water vapor

Curve 681141-A

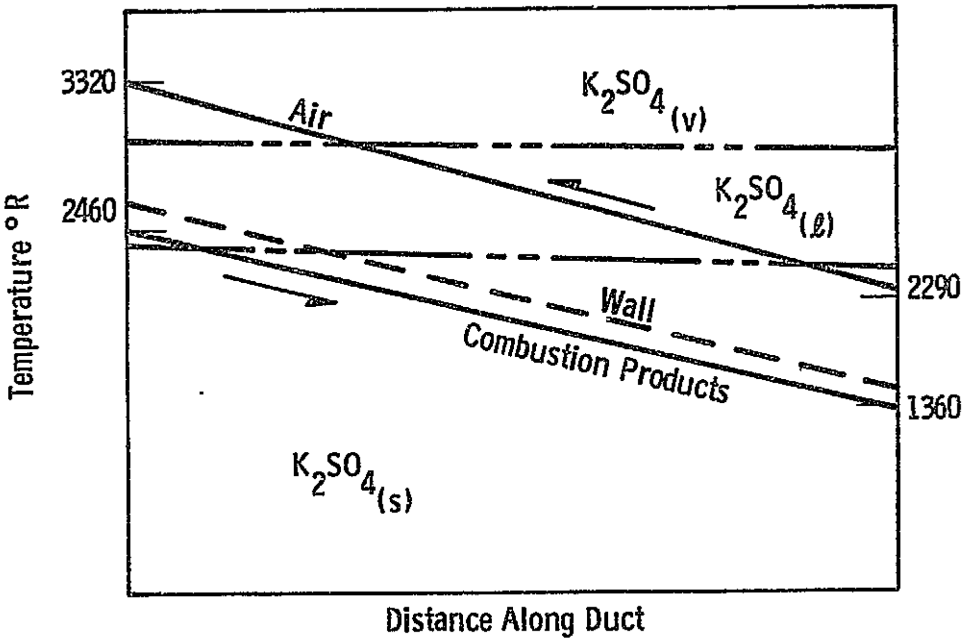

Fig. A 9.2.6—Temperature distance chart for the recuperative heat exchanger, metalic section

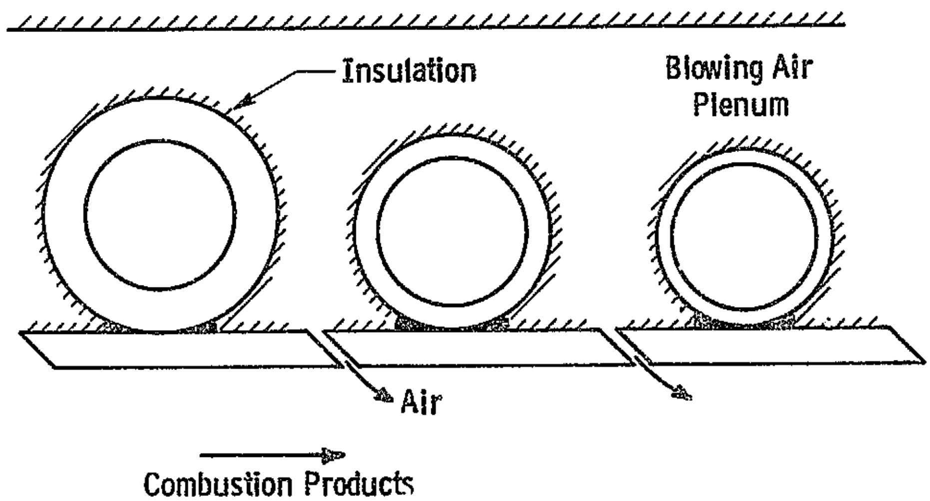

Fig. A 9.2.7—Cross section of metalic recuperator tubes

and carbon monoxide, respectively, by techniques previously documented. For these conditions the emmissivities are computed as:

$$\varepsilon_{g_{2807}} = \varepsilon_{g_{CO_2}} + \varepsilon_{g_{H_2O}} - \Delta\varepsilon = 0.153 + 0.21 = 0.05 = 0.313 \quad \text{(A 9.2.43)}$$

$$\varepsilon_{g_{2060}} = 0.386 \quad \text{(A 9.2.44)}$$

A radiation heat exchange coefficient, h_r as defined by Equation A 9.2.3 is found to be 162.4 W/m^2-°K (28.6 Btu/hr-ft^2-°F).

The combined coefficient, U_{r+c}, is assumed to be 141.9 W/m^2-°K (25.0 Btu/hr-ft^2-°F) when corrected for blowing the boundary layer as in Equation A 9.2.37. Figure A 9.2.7 shows a cross section of the proposed wall tube.

Conductance per foot of tube is given by Equation A 9.2.45:

$$(UA_f)' = \frac{1}{\frac{1}{\eta_t \pi d_i h_i} + \frac{\ell n(d_o/d_i)}{2\pi\ k_t} + \frac{1}{\eta_f\ A_f{}'\ U_{c+r}}} \quad \text{(A 9.2.45)}$$

For a 1.27 cm (0.5 in) thick by 10.2 cm (4 in) wide flange, fin efficiency, η_f, is given for a rectangular fin as:

$$\eta_f = \frac{1}{m\ell} \tanh\ (m\ell) \quad \text{(A 9.2.46)}$$

where ℓ in this case is half the flange width. Substitution of average material properties yields the value of $m\ell$ given by Equation A 9.2.47.

$$m\ell = \sqrt{\frac{U_{r+c}}{K}\ \frac{1}{t}}\ \ell = \sqrt{\left(\frac{25}{11.5}\right)\left(\frac{12}{0.5}\right)}\ \frac{1.75}{12} = 1.05 \quad \text{(A 9.2.47)}$$

Substitution into Equation A 9.2.46 yield an effectivness, η_f of 0.75. A similar value is used to account, approximately, for varying tube wall temperature.

Substitution of the following parameters in Table 9.2.6 into Equation A 9.2.18 gives a tube-side convection coefficient, h_i, of 368.7 W/m^2-°K(65 Btu/hr-ft^2-°F).

Table A 9.2.6 Tube-Side Air Properties in Metallic Recuperator

Parameter	Value
$Pr^{1/3}$	1.0
d_i	1/6 ft
$\bar{k}_a$	0.0477 Btu/hr-ft-°F
ρ	0.131 lb/ft^3
μ	3.33 x 10^{-5} lb/ft-s
V	150 ft/s
Re_{150}	0.983 x 10^5
η_t	0.75
A_o'	1/3 $ft^2_{H.X.}/ft_{tube}$

Values also substituted into Equation A 9.2.45 from Table 9.2.6 yield an overall conductance per unit length of tube, (UA)' of 8.306 W/m^2-°K (4.8 Btu/hr-ft-°F).

To heat 1256 kg/s (2769 lb/s) of air from 756 to 1367°K (900 to 2000°F) requires the addition of 882.2 MJ/s (3.01×10^9 Btu/hr) of heat. With a log mean temperature difference of 498°K (897°F) and the previously cited unit conductance approximately 213,133 m (699,250 ft) of tube are required. This solution is, of course, iterative with only the last iteration presented. The length of the tube is equivalent to 21,654 m^2 (233,000 ft^2) of irradiated surface. With four vertical splitters in a 15.85 m (52 ft) square duct with a slag tap floor, the metal section of the exhaust duct becomes 129 m (423 ft) long.

Circuitry must be very nearly true counterflow. This can be accomplished by 4 tube-side passes, 7000 parallel tubes per pass. The configuration has a relatively high pressure drop of 34.47 kPa (5 psi) and should be optimized for pressure drop and cross-flow correction factor in Task II.

A 9.2.6.1 Metallic Recuperator Scaling

Several characteristics make scaling of the metallic recuperator from case to case comparatively simple. As in the case of the ceramic recuperator, the design is sufficiently flexible to be applicable to all cases. The metallic recuperator always heats air to the same top temperature, 1367°K (2000°F), and takes air from the diffuser cooler. It should be noted that although the design is very flexible, the duct wall dimension is proportional to the square root of mass flow. It is assumed that air velocity and tube length can be adjusted over a range of cases so that tube-side thermal impedance variation remain negligible. Under these conditions, the heat exchange surface area is directly proportional to the heat added to the air and inversely proportional to the mean temperature between the exhaust products and the air, as shown in Equation A 9.2.4 where the temperatures are in °R.

$$A_s \propto \frac{\dot{m}_a \,(2460 - T_{a_i})}{T_{p_i} - 2460 + T_{p_o} - 1360} \qquad \text{(A 9.2.48}$$

The constant of proportionality is obtained from the case computed herein.

Table A 9.2.7 - Metallic Recuperator Material

Metallic Tube Weight	48.7 $lb/ft^2_{H.X.}$ (34.3 $lb/ft^2_{H.X.}$ for lower mass flows)
Headers (10% tube)	4.87 $lb/ft^2_{H.X.}$
Refractory Insulation (1)(1)(1) ft^3 (43) lb/ft^3	43 $lb/ft^2_{H.X.}$

Table A 9.2.8 - Recuperator Structural Materials

Concrete Slab (1.5 ft thick)	1624 yd^3
Structural Steel:	
12 x 3 in channel arch every 15 ft (34 arches) 34 x 150 x 25 lb/ft	128,000 lb
Cross Framing	
34 x 50 x 25 lb/ft	43,000 lb
Shell Plate, 1/2 in	20.4 $lb/ft^2_{H.X.}$

A 9.2.6.2 Recuperator Operating and Maintenance

The recuperator would probably be cleaned by sand blasting once a year, and 2.0% of the surface should require patch welding each year. Costs of these operations are given in Table A 9.2.9

Table A 9.2.9 Annual O&M Charges

Operation	Cost Material & Labor, $/ft^2$	Direct Charge, $/ft^2$-yr
Sand Blasting	1.50	1.50
Welding	3.30	6.60

A 9.2.7 Separately Fired Air Heater

Periodic air preheaters of the blast furnace stove-type are recommended for relatively clean fuel gas. High-temperature gasified coal fuel gas can probably be cleaned to conditions of 0.57 g/m^3 (0.25 gr/ft^3) dust loading and 800 ppm combined alkali and chlorine. Under these circumstances a reasonable life [157.7 to 220.7 Ms (5 to 7 yr] may be obtained from refractory checkers. Although various techniques (high-pressure burners and air-pressure drop, close approach to heating gas with nearly constant temperature checkers, etc.) may be used to increase the heat rate per unit checker mass, all of these techniques increase the refractory duty and incur their own cost penalties. For these reasons, a conventional model of a regenerator is used.

In this model the center line of a checker at a given elevation is assumed to remain at constant temperature through a heat-cool cycle. The checker surface temperature fluctuates according to the periodic temperature solution given by Carslaw and Jeager (Reference 9.33). If the checker center-line temperature remains constant, then the heat flow may be approximated by that of a recuperative type, in other words, linear temperature profiles. It is evident that fouling is twice as significant here as it is in a recuperative heat exchanger

with clean and dirty streams since it represents a thermal impedance to heat entering and leaving the checkers. The same may be said for pressure drop due to fouling or slump.

Initial sizing of a stove may be obtained from Equation A 9.2.40, using the conductance per foot given by Equation A 9.2.49 (Reference 9.25)* for symmetrical heating and cooling cycles. The source of the resistances is readily apparent except for the θ term, an equivalent resistance of heat storage which in effect approximately accounts for nonlinear temperature gradients in the checker. This term is neglected for all initial sizing, being only 2% of the overall thermal impedance. Fouling is accounted for as an additional series resistance with convection. In dusty environments Trinks (Reference 9.34) mentions as much as 20% reduction in heat exchange coefficients. Figure 1 of Reference 9.35 suggests a 25% augmentation of the convection coefficient for straight flues, to give the convection coefficient for flues made from 12.06 cm (4.75 in) high brick and 3.18 mm (0.125 in) misalignment in stacking. Spoiler effects are considered to exactly offset fouling effects.

$$(UA)' = \frac{1}{\dfrac{1}{h_g A_s'} + \dfrac{r_B}{k A_s'} + \dfrac{1}{h_a A_s'} + \dfrac{\theta}{2.5\ C_s \rho_s r_B A_s'} + \dfrac{2}{h_{f1} A_s'}} \qquad \text{(A 9.2.49)}$$

The characteristic checker dimension, r_B, is defined by Equation A 9.2.50.

$$r_B = \frac{\Psi_s}{A_s} \qquad \text{(A 9.2.50)}$$

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*Note: McAdams's Equation 11-35 uses a term $r_B/[k(\theta''+\theta')]$ which should read $2\ r_B/[k(\theta''+\theta')]$.

Flue surfaces in these designs are rough square tubes, formed by straight, noninterconnected chimmney flues, similar to basketweave checker settings. The void volume, ε, is given as:

$$\varepsilon = 1 - \frac{\mathcal{V}_s}{\mathcal{V}} \qquad \text{(A 9.2.51)}$$

McAdams (Reference 9.25) gives the regenerator equations developed by Hansen and Hottel as the following, Equations A 9.2.52 through A 9.2.61.

Dimensionless regenerator size:

$$\lambda_g = \frac{h_g}{C_{p_g} G_{o_g}} \left(\frac{1 - \varepsilon}{r_B}\right) L \qquad \text{(A 9.2.52)}$$

$$\lambda_a = \frac{h_a}{C_{p_a} G_{o_a}} \left(\frac{1 - \varepsilon}{r_B}\right) L \qquad \text{(A 9.2.53)}$$

Dimensionless regenerator period as:

$$\tau_g = \frac{h_g \theta}{C_s \rho_s r_B} \qquad \text{(A 9.2.54)}$$

$$\tau_a = \frac{h_a \theta}{C_s \rho_s r_B} \qquad \text{(A 9.2.55)}$$

Steam rise or fall:

$$\frac{\left(T_1 - T_2\right|_g}{\Delta T_{\ell m}} = \frac{\left(\lambda/\tau\right|_g}{\tau_g^{-1} + \tau_a^{-1} + 2\left(\tau_g + \tau_a\right)^{-1} \left[\lambda_s \left(\frac{1}{\phi_s} - 1\right) - 2\right]} \qquad \text{(A 9.2.56)}$$

$$\frac{\left(T_1 - T_2\right)_a}{\Delta T_{\ell m}} = \frac{\left(\lambda/\tau\right)_a}{\tau_g^{-1} + \tau_a^{-1} + 2\left(\tau_g + \tau_a\right)^{-1}\left[\lambda_s\left(\frac{1}{\phi_s} - 1\right) - 2\right]} \qquad \text{(A 9.2.57)}$$

The mean effectiveness, ϕ_s, is given in Figure 11-11 of Reference 9.25 using a mean size and period of

$$\lambda_s = 2\left(\lambda_g^{-1} + \lambda_a^{-1}\right)^{-1} \qquad \text{(A 9.2.58)}$$

and

$$\tau_s = \left(\tau_g + \tau_a\right) 1/2 \qquad \text{(A 9.2.59)}$$

In the regenerator design, so far, mean portal temperatures are assumed. The thermal droop is approximated as:

$$\frac{\delta T_{g_2}}{T_{g_2} - T_{a_1}} = \frac{\tau_g}{\left(\left(\phi_s^{-1} - 1\right)\left(\lambda_g + 1\right)\right)} \qquad \text{(A 9.2.60)}$$

and

$$\frac{\delta T_{a_2}}{T_{g_1} - T_{a_2}} = \frac{\tau_a}{\left(\phi_s^{-1} - 1\right)\left(\lambda_a + 1\right)} \qquad \text{(A 9.2.61)}$$

With the plurality of design variables represented in Equation A 9.2.49 through A 9.2.61, it is evident that the appoximation by Equations A 9.2.49 and A 9.2.40 and temperature difference is very useful. Given the constraints of solid chimmney flues and using minimum dimensions consistent with slag or dust loading places constraints upon the design. Consistent design parameters are achieved by iteration.

A uniform set of stove materials is used in all ECAS stove-type regenerators. In keeping with industry practice, refractories are used to within 100°K (180°F) of their maximum environmental service temperature. Table A 9.2.1 indicated the thermal ranges of materials. In any computation weighted average cost and properties are used for the whole stove.

The introduction contains guidelines from industry practice on flue size versus dust loading. Because stove efficiency and low cost both favor small flue size, and because historically refractory manufacturers have increased checker duty with increasing raw material purity, the industrial guidelines are eased in this design to allow 5.08 cm (2 in) square holes with 0.572 g/m^3 (0.25 gr/ft^3) dust loading. In reality this does not lie far outside the guidelines, because industrial guidelines are based on low-Btu gas, and the Westinghouse proposed gasifiers produce low- to medium-Btu gas. For fuel air ratios giving 10% excess air the combustion product dust loadings are comparable.

The final iteration of the Base Case 1, Point 1, stove design is given as an example. An additional constraint on the design is that cycle time should be short for low capital cost and long compared to blowdown and valve cycle times. Modern automated blowdown and valve operation stove change systems yield 180 to 300 s (3 to 5 min) turn around. Thus, the stove time period should not be less than 600 s (10 min).

Using the 600 s (10 min) cycle in the mean Fourier Number and mean Biot Number, the Heisler Charts indicate a 3.18 cm (1.25 in) thick wall will have a center-line temperature change of about $\pm$ 10%. This is a reasonable range for designs not using the saturated checker approach being operated in parallel to avoid thermal droop. For an 8.255 cm (3.25 in) square brick with a 5.08 cm (2 in) square flue, the void fraction is 0.379, and the characteristic dimension, r_B, is 0.02083 m (0.068 ft). The ratio of gas to air velocity, assuming negligible heat loss rate, is given by Equation A 9.2.62.

$$\frac{V_g}{V_a} = \frac{\rho_a \; A_s \; C_{p_a} \left(T_{2_a} - T_{1_a} \right) \theta_a}{\rho_g \; A_s \; C_{p_g} \left(T_{1_g} - T_{2_g} \right) \theta_g} \tag{A 9.2.62}$$

Table A 9.2.10 is a compendium of various parameters applicable to stove operating conditions.

Table A 9.2.10 Stove Parameters

Parameter	Value
$\dot{m}_a$	1532 kg/s
$\dot{m}_p$	790 kg/s
ρ_a	0.0855 lb/ft^3
ρ_p	0.0099 lb/ft^3
ρ_s	173 lb/ft^3
μ_a	4.02×10^{-5} lb/ft-s
μ_p	4.43×10^{-5} lb/ft-s
$Pr^{1/3}_{a,p}$	1.0
V_a	30 ft/s
V_p	134 ft/s
d_h	1/6 ft
Re_a	0.107×10^5
Re_p	0.0497×10^5
$\bar{k}_s$	1.31 Btu/hr-ft-°F
k_a	0.0589 Btu/hr-ft-°F
k_p	0.0652 Btu/hr-ft-°F
C_{p_s}	0.264 Btu/lb-°F
C_{p_a}	0.312 Btu/lb-°F
C_{pp}	0.339 Btu/lb°F

Substitution of values from Table A 9.2.10 into Equation A 9.2.18 yields the heat exchange coefficient for products and air which are 45.99 and 76.64 W/m^2-°K (8.1 and 13.5 Btu/hr-ft^2°F), respectively.

For an initial estimate of the stove size, the value of r_B used is the checker wall thickness. This is consistent with the steady recuperator model of the regenerator. Substituting into Eqation A 9.2.49 while neglecting heat storage impedance yields the conductance per unit length of flue:

$$(UA)' = \left(\frac{12}{13.5(8)} + \frac{1.25}{1.31(8)} + \frac{12}{8.1(8)} \right)^{-1} = 2.41 \text{ Btu/hr-ft-°F} \quad \text{(A 9.2.63)}$$

The active length of a flue in this signal pass regenerator is:

$$L = \frac{\overline{V} A_f \rho \, \overline{C}_p \, \Delta t_a}{(UA)' \, \Delta T_{\ell m}} \quad \text{(A 9.2.64)}$$

Substitution of temperature from Figure A 9.2.8 into the usual equation for counterflow log mean temperature difference yields:

$$\Delta T_{\ell m} = \frac{4060 - 3400 - (3060 - 2842)}{\ln \left(\frac{4060 - 3400}{3060 - 2842} \right)} = 400\text{°F} \quad \text{(A 9.2.65)}$$

Substituting the temperature rise of air from Figure A 9.2.8, the $\Delta T_{\ell m}$ and (UA)', and values from Table A 9.2.10 into Equation A 9.2.64 yields:

$$L = \frac{(30)(3600)(4/144)(0.0855)(0.294)(3400 - 2842)}{2.41(400)} = 43.69 \text{ ft} \quad \text{(A 9.2.66)}$$

Subsequent iteration shows that the flue height only needs to be 12.80 m (42.0 ft) to obtain the required air-stream temperature rise.

Curve 681140-A

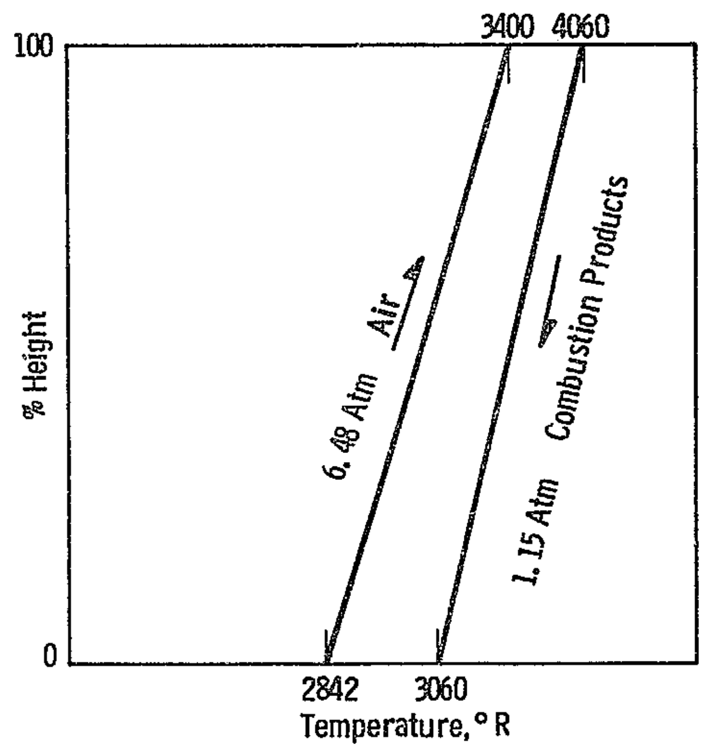

Fig. A 9.2.8—Temperature-height chart for the separately fired air heater

Substituting values for a 600 s (10 min) cycle into Equations A 9.2.52 through A 9.2.59 yield:

$$\lambda_a = \frac{13.5}{0.312(0.0855)(30)(0.379)} \left(\frac{1-0.379}{0.0684}\right) \frac{42.1}{3600[s/hr]} = 4.73 \qquad \text{(A 9.2.67)}$$

$$\lambda_g = \frac{8.1}{0.339(0.0099)(133.5)(0.379)} \left(\frac{1-0.379}{0.0684}\right) \frac{(42.1)}{3600} = 5.075 \qquad \text{(A 9.2.68)}$$

$$\tau_a = \frac{13.5\ (1/6)}{(0.264)(173)(0.0684} = 0.72 \qquad \text{(A 9.2.69)}$$

$$\tau_g = \tau_a\, h_g/h_a = 0.432 \qquad \text{(A 9.2.70)}$$

$$\lambda_s = 4.90 \qquad \text{(A 9.2.71)}$$

$$\tau_s = 0.576 \qquad \text{(A 9.2.72)}$$

From Figure 11-11 of Reference 9.25, the mean effectiveness, ϕ_s, is found to be 0.7.

Substituting into Equation A 9.2.57 for air temperature rise gives:

$$(T_2 - T_1)_a = \frac{(4.73/0.72)\ (400)}{2.31 + 1.39 + 1.74\ [4.9\ (\frac{1}{0.7} - 1) - 2]} = 679°F \qquad \text{(A 9.2.73)}$$

The air preheat temperature is 1956°K (3061°F), which exceeds the 1889°K (2940°F) required.

The temperature drop of the gas products stream is given by Equation A 9.2.56 or:

$$T_1 - T_2\big|_g = 679\ (5.08/0.432)\ (0.72/4.73) \approx 1215°F \qquad \text{(A 9.2.74)}$$

This compares favorably with the 556°K (1000°F) required. The air thermal droop calculated using Equation A 9.2.61 is less than 111°K (200°F). Since the calculated air rise is 67°K (121°F) greater than required, this will partially offset the thermal droop, or a shorter time cycle may be used. This design is probably not optimum but is workable. The basic stove structural design with combustion chamber is typical and along the lines of Figures 110a and 110b of Reference 9.?3. Note that the combustion chamber and perimeter insulation provides reserve capacity. This capacity is greater than usual because the stove shell is designed to a 6.1 m (20 ft) diameter to facilitate section arches under the checkwork.

The final wrap-up on this stove design requires the superficial velocity, V_o, given as

$$V_o = \varepsilon\, V_a = 0.379(30) = 11.4 \text{ ft/s} \tag{A 9.2.75}$$

Total active checker flow area is given by the continuity equation expressed as Equation A 9.2.76:

$$A = \frac{\dot{m}_a}{V_o\, \rho_a} = \frac{1532[\text{kg/s}]\ 2.2[\text{lb/kg}]}{11.4[\text{ft/s}]\ 0.0955[\text{lb/ft3}]} = 3465 \text{ ft}^2 \tag{A 9.2.76}$$

Each stove has an active checker superficial area of 16.5 m^2 (177 ft^2). Therefore, 40 stoves are required, 20 being heated while 20 others are giving up their heat to the air. The stove pressure drop is small (approximately 2.07 kPa (0.3 psi).

A 9.2.7.1 Stove Regenerator Scaling

The stoves are scaled in the same way as are recuperators, implying, as did the assumption of negligible heat storage impedance, that heat storage will not become a problem. Since flue velocity and wall thickness are constant, this is reasonable. Unfortunately, primary combustion air preheat temperatures diminish in some cases

to a point where the capital cost of a stove is not justified, and some other means of preheating air is indicated (a muffle heater, for example). Separate designs were not justified because as the stove height diminishes [42 stoves 1.83 m (6 ft) high for Base Case 1, Point 8, borders on the ludicrous] the stove cost becomes a small part of plant cost.

The scaling rule is again written as a proportion, the constant being determined by the above computation of stove size for Base Case 1.

The height of the stove was assumed to be directly proportional to the required air temperature rise and inversely proportional to the stove log mean temperature difference:

$$L \propto \Delta T_a / \Delta T_{\ell m} \qquad \text{(A 9.2.77)}$$

The number of stoves was assumed to be proportional to the mass flow rate of the air.

$$N \propto \dot{m}_a \qquad \text{(A 9.2.78)}$$

A 9.2.7.2 Stove Regenerator Material

Traditional stove design incorporates a combustion chamber and active checkerwork in a refractory lined shell. In this case the 1.905 cm (0.75 in) steel shell is lined with 0.457 m (1.5 ft) of refractory brick insulation. The combustion chamber and required insulation is assumed to occupy 22% of the available internal cross section. On the heat cycle, preheated air and fuel are burned, flowing upward in the combustion chamber. The hot mixture flows out of the chamber, over the dome, and downward through the heat storage checkers. Although a traditional single down-flow pass has been used, parallel multipass circuiting of the units is just as feasible as for high-temperature recuperators. Such optimization should be continued in Task II.

For the above units, the active materials are given as follows:

Refractory brick insulation = (thickness(circumference)(height)(density)

$$= (1.9)(20\pi)(42 + 10)(80) = 496{,}622 \text{ lb/stove} \quad (A\ 9.2.79)$$

Checker brick = (height)(free area) $(1 - \epsilon)$(density)

$$= (42)(177)(1 - 0.379)(200) = 923{,}300 \text{ lb/stove} \quad (A\ 9.2.80)$$

Table A 9.2.11 - Stove Regenerator Structural Materials

Concrete (reinforced)		
Footer	18 ft id by 22 ft od by 6 ft thick	28 yd^3
Slab	18 ft dia. by 1.5 ft thick	14 yd^3
Steel Plate Shell		
$\pi\ 20(42.1) + 2\pi(10)^2 = 3272\ ft^2$		
$3272\ [ft^2]\ 30.6\ lb/ft^2 = 10^5\ lb$		

A 9.2.7.3 Stove Regenerator Operating and Maintenance Costs

The top 10% of the checker brick will probably need to be replaced each year.

A 9.2.8 Separately Fired Air Heater Combustion Air Preheater (CAP)

The function of the CAP is to preheat the stove combustion air. Products of combustion from the stove at 1700°K (2600°F) (see Figure A 9.2.8)

are divided into two streams: one recirculates to the stove, and the other flows through the CAP. Heat is recovered from the products stream [335 kg/s 2658 x 10^6 lb/hr)] by the air stream [284 kg/s (225 x 10^6 lb/hr)] a counterflow recuperator (see Figure A 9.2.9) of the muffle type (Figure A 9.2.10). The ducts are assumed to have a heat exchange coefficient of 85.2 W/m^2-°K (15 Btu/hr-ft^2-°F) on both sides. Areas are assumed equal, and the overall conductance per unit area, U, is given by the usual relation as:

$$U = \frac{1}{\frac{1}{h_a} + \frac{1}{h_p} + \frac{t}{k_s}} \qquad \text{(A 9.2.81)}$$

$$U = \left[\frac{1}{15} + \frac{(\frac{1}{12})}{1.3} + \frac{1}{15}\right]^{-1}$$

$$= [1/15 + 1/[(1.3)(12)] + 1/15]$$

$$= 5.07 \text{ Btu/hr-ft}^2\text{-°F}$$

Log mean temperature difference is given as Equation A 9.2.82 using temperatures from Figure A 9.2.9:

$$\Delta T_{\ell n} = \frac{3060 - 2860 - (1110 - 520)}{\ell n(\frac{3060 - 2860}{1110 - 520})} = 360\text{°F} \qquad \text{(A 9.2.82)}$$

The required surface area is given by Equation A 9.2.83:

$$A_s = \frac{(\dot{m}_a)(C_{pa}(\Delta T_a)}{(U)(\Delta T_{\ell n})} \qquad \text{(A 9.2.83)}$$

$$= \frac{2.2(284)(0.2815)(2340)\ 3600}{5.07(360)} = 8.12 \times 10^5 \text{ft}^2$$

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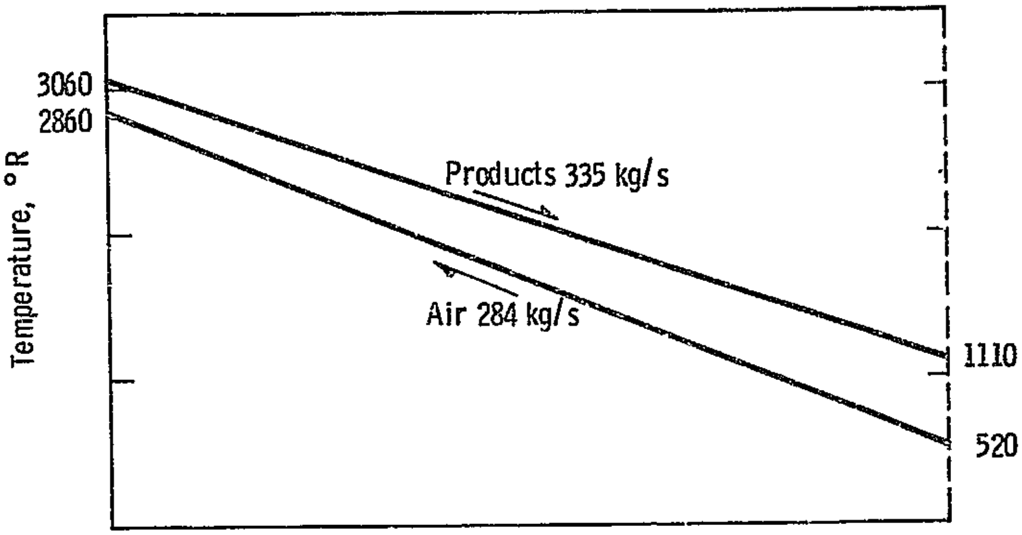

Fig. A 9. 2.9 —Temperature distance chart for the separately fired air preheater

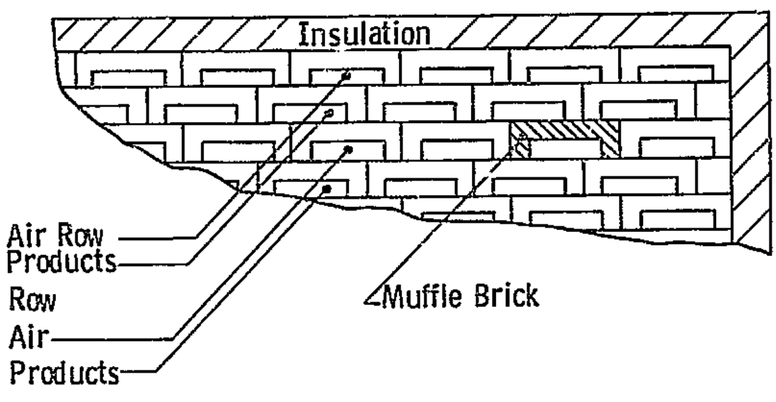

Fig. A 9. 2. 10—Cross section of a typical muffle type air heater

Velocity (mean of products and air) is assumed to be 15.2 m/s (50 ft/s). Approximate flow area is computed as:

$$A_f = \frac{(300)(2.2)}{(50)(0.0169)} = 781 \text{ ft}^2 \qquad \text{(A 9.2.84)}$$

For a 7.62 by 17.78 by 5.715 cm (3 x 7 by 2.25 in) muffle brick with a 12.7 by 5.08 cm (5 by 2 in) channel, the number of channels per stream is:

$$N = \frac{(781)(144)}{(5)(2)} = 11{,}250 \qquad \text{(A 9.2.85)}$$

The height is computed as:

$$L = \frac{(0.812 \times 10^6)(12)}{(0.01125 \times 10^6)(14)} = 61.6 \text{ ft} \qquad \text{(A 9.2.86)}$$

Based on the specified brick, the total area of the muffle is:

$$\text{Muffle area} = 2(11{,}250)\left(\frac{21}{144} \text{ ft}^2\right) = 3280 \text{ ft}^2 \qquad \text{(A 9.2.87)}$$

For a square unit one side is 17.5 m (53.3 ft).

A 9.2.8.1 CAP Scaling

The CAP is scaled in a manner similar to that used for other apparatus, such as the stoves. Unfortunately, the CAP computation above

did not correspond to any base case or variation and had to be scaled even for the base case.

A 9.2.8.2 CAP Materials

For Base Case 1. Point 1, the superficial flow area given by Equation A 9.2.18 is multiplied by 0.841 because of fuel/air ratio and flow rate scaling, and the height by 0.482 to account for temperature and heat rate scaling. The muffle brick volume is then:

$$\text{Brick Volume} = (0.841)(0.482)(61.6)(3280) = 81{,}400 \text{ ft}^3 \qquad \text{(A 9.2.88)}$$

A superficial muffle brick density of 320 kg/m^3 (20 lb/ft^3) is assumed to yield 748 Mg (825 tons) of muffle brick. The muffle brick weight is summed with the SFAH checkers and must be appropriately weighted for cost differential, as in Table A 9.2.11.

Table A 9.2.11 - Muffle Brick Average Δ-T

Cost of SFAH check	\$436/1000 lb	(Table 9-22)
Cost of CAP Muffle Brick	\$233/1000 lb	
Muffle Brick Equivalent Weight	1550 (233/(436) = \$880/1000 lb	

A 9.2.8.3 CAP Insulation

The insulation on the CAP is assumed to be on average 0.203 m (8 in) thick. The volume is given as:

$$\left(\frac{2}{3}\right)[4(0.482)(71.6)\sqrt{(0.841)(3260)} + 3260] = 6330 \text{ ft}^2 \qquad \text{(A 9.2.89)}$$

Using 400 kg/m^3 (25 lb/ft^3) insulation yields a weight of 71.7 Mg (79 tons) of insulation, which is also weighted by SFAH insulation cost.

CAP Insulation Weight = 158 (210)/230 = 145,000 lb (A 9.2.90)

Table A 9.2.12 - CAP Structural Materials

Reinforced Concrete Footer	
$4(52.2)(5)(2)/27 = 77\ yd^3$	
Slab	
$(1.5)(50)(50/27) = 140\ yd^3$	
Plate	79,000 lb
Structural Steel	16,000 lb

A 9.2.9 Sample Tabular Display of Data

The following section demonstrates how the data of Appendix A 9.2 are displayed in the tabular data of Tables 9.10 and 9.22.

In Table 9.10, the amount in Subaccount 13.1 entitled Ceramic Tubing is the sum of all ceramic tubing weight in the high-temperature recuperator.

The product of the panel length, panel height, the number of panels, and the weight per unit of heat transfer surface gives the weight of ceramic tube:

Weight ceramic tube = (82)(52)(11)(13.4) = 635,000 lb (317.5 tons) (A 9.2.91)

In Table 9.10, the amount in Subaccount 13.2, entitled Exotic Metal Tubes, is the sum of all material specified as RA333. Half of the metallic recuperator is assumed to be made from RA333 and half from corrosion-resisting stainless, Type 347. Headers for the RA333 are refractory lined and do not enter here. Note that the metallic recuperator

length given in Section A 9.2.6 is 129 m (423 ft). The length undergoes scaling changes to 130 m (427 ft) because the diffuser heat rate scales product inlet temperature downward. The amount in Subaccount 13.2 is computed as the product of one-half the panel length, the panel width, number of panels, and metallic tube weight unit air of heat transfer surface.

$$\text{Weight of refractory tube} = \frac{1}{2}\,(427)(52)(11)(48.7) = 5{,}810{,}000 \text{ lb} = 2{,}905 \text{ tons} \tag{A 9.2.92}$$

In Table 9.10, the amount in Subaccount 13.3, entitled Stainless Steel Tubes, includes diffuser tubes and the other half of the recuperator tubes [2.635 Gg(5,810,000 lb)].

The diffuser tubing weight is computed as:

$$\text{Diffuser weight} = 4(\text{mean width})(\text{length})(\text{weight/ft}^2 \text{ surface})$$

$$= 4\left(\frac{12.8 + 52}{2}\right)(147)(25.3) = 483{,}000 \text{ lb} = 241.5 \text{ tons} \tag{A 9.2.93}$$

Note that the axial length 44.8 m (147 ft) is used interchangeably with the linear wall dimension in this estimate and that 25.3 is the sum of the flange and tube weight per square foot of surface area. The total weight of stainless steel is, therefore, 2.855 Gg (6,293,000 lb or 3146.5 tons).

In Table 9.10, the amount in Subaccount 13.4, entitled Tube Ceramic Coating, concerns the ceramic coating on the diffuser tubes. This is calculated as the product of the diffuser surface area and the weight per unit area of the deposited material:

$$\text{Weight of ceramic coating} = 4\left(\frac{12.8 + 52}{2}\right)(147)(8.2) = 157{,}000 \text{ lb} = 780.5 \text{ tons} \tag{A 9.2.94}$$

In Table 9.10, the amount in Subaccount 13.5, entitled Insulation for Regenerator, is the sum of refractory insulation for the diffuser and recuperator. The weight of insulation in the diffuser is approximated by assuming a diffuser surface area of 1858 m^2 (20,000 ft^2) and a material weight per unit surface area of 688.8 kg/m^3 (43 lb/ft^3).

$$\text{Diffuser insulation weight} = (20,000)(43) = 860,000 \text{ lb} = 430 \text{ tons} \quad (A\ 9.2.95)$$

Similarly, for an assumed recuperator area of 6968 m^2 (75,000 ft^2), the insulation weight would be 1.463 Gg (3,226,000 lb). The total weight of insulation would be the sum [1.853 Gg (4,086,000 lb)].

In Table 9.10, the amount in Subaccount 13.6, entitled Structural Steel, is the sum of diffuser and regenerator material. Structural steel entries in Tables A 9.2.3 of 9.07 Mg (20,000 lb) and entries in Table A 9.2.8 of 58.06 and 19.5 Mg (128,000 and 43,000 lb) give a sum of 86.64 Mg (191,000 lb).

In Table 9.10, the amount in Subaccount 13.7, entitled Containment Steel Regenerator, the containment steel being a 1.27 cm (1/2 in) thick sheet, is computed as Item 13.5:

$$\text{Containment Steel Weight} = (95,000 \text{ ft}^2)\ (20.4 \text{ lb/ft}^2) = 1,938,000 \text{ lb}$$

In Table 9.10, Subaccount 13.22 entitled, Concrete (Reinforced), item 13.22 is the sum of diffuser and recuperator slab and footer material. From Tables A 9.2.3 and A 9.2.8, the quantity of reinforced concrete is:

$$\text{Volume Concrete} = 279 + 1624 = 1903 \text{ yd}^3$$

In Table 9.10, Subaccount 13.20, entitled Headers, header weights are estimated as:

- Diffuser 48,000 lb
- Ceramic recuperator 254,000 lb
- Metallic recuperator 343,000 lb

In Table 9.22, Subaccount 13.8 entitled, Checker Bricks, the weight of checker bricks includes checker bricks and muffle bricks. The 17.27 Gg(38,080 klb) is the sum of 0.399 Gg(880 klb) of muffle brick (Table A 9.2.11) and 16.99 Gg(37,200 klb) of SFAH checkers. SFAH checkers are computed as the product of 9.3×10^5 lb/unit from Equation A 9.2. and 40 units.

In Table 9.22, Subaccount 13.9, entitled Insulation for SFAH, the SFAH insulation also includes CAP refractory brick insulation.

$$\text{Insulation Weight} = (40)(4.966 \times 10^5) + 145,000 = 20,145,000 \text{ lb}$$

In Table 9.22, Subaccount 13.10, entitled Containment Steel, Item 13.10, SFA heater containment steel, consists of 1.814 Gg (4,000,000 lb) of SFAH shell and 33.6 Mg (74,000 lb) of CAP shell (Table A 9.2.12).

In Table 9.22, Subaccount 13.11, entitled Structural Steel, Item 13.11, consists of 0.381 Mg (840,000 lb) of SFAH structural steel and 7.25 Mg (16,000 lb) of CAP structural steel.

A 9.2.10 Nomenclature

A = Area

A_s = Surface area, wetted, irradiated

C_L = Fluid energy loss coefficient

C_p = Constant pressure specific heat

C_s = Specific heat solid checker

d = Diameter

f = Friction factor, Darcy-Weisbach

G_o = Superficial mass velocity

g_c = Gravitation constant

h_c = Heat convection coefficient

h_D = Mass diffusion coefficient

h_r = Heat radiation coefficient

i = Specific enthalpy

k = Thermal conductivity

L = Length

L_o = Equivalent radiation beam length, Equation A 9.2

ℓ = Length

$\dot{m}$ = Mass flow rate

m = Square root of the thermal impedance ratio, internal conduction/ surface convection

P = Pressure

P_p = Partial pressure

P_pL = Radiation opacity term, ft-atm

Pr = Prandtl number

$\dot{q}$ = Heat rate

Re = Reynolds number

r_B = Characteristic checker thickness, Equation A 9.2.19

S = Length of side

T = Absolute temperature

$\Delta T_{\ell m}$ = Log mean temperature difference, counterflow

t = Thickness

U = Combined heat exchange coefficient

V = Velocity

V_o = Superficial velocity

$\mathcal{V}$ = Volume.

Greek

α = Absorbtivity

Δ = Incremental change

$\Delta\varepsilon$ = CO_2 - H_2O radiation interaction correction

δ = Incremental change

ε = Emissivity, void fraction

η = Fin efficiency

Θ = Time period

λ = Nondimensional stove size

μ = Absolute viscosity

ρ = Density

σ = Stephan-Boltzmann Constant, working stress

τ = Nondimensional stove time period

ϕ = Heat exchange effectiveness, effective

Subscripts

a = Air

c = Convection

f = Fin, flow

fl - Fouling

g, G = Gas

i = Inlet, inside

L = Length

r = Radiation

s = Solid, surface

t = Tube, thickness

1 = Inlet, inside

2 = Outlet, outside.

Superscripts

' = Per unit length

" = Per unit area.

Appendix A 9.3

COUPLING HEAT EXCHANGER

A 9.3.1 Description of the Duty of the Coupling Heat Exchanger

The steam generator for the open-cycle MHD system transfers heat from potassium-seeded combustion products (the working fluid of the MHD cycle) to water (the working fluid of the steam bottoming cycle). For this reason, the steam generator is known as the coupling heat exchanger.

The description which follows applies specifically to the steam generator for Base Case 1, and methods used to scale the design and costs to satisfy other base cases will be described in Section A 9.3.6.

In Base Case 1, 1449.2 kg/s (1.15×10^7 lb/hr) of seeded combustion products enter the steam generator at 1610°K (2438°F). This primary stream must leave the steam generator at 425°K (306°F) for passage to the main seed-removal facility. This represents a heat transfer rate of 2166 MWt. On the secondary side, the feedwater must be raised from 402°K (266°F) to steam at 811°K (1000°F), the throttle pressure being 24.13 MPa (3500 psi). Additionally, the full flow of steam should be reheated within the steam generator from 664°K (735°F) to 811°K (1000°F) at 7.584 MPa (1100 psi). Within these constraints the steam flow is 693 kg/s (5.5×10^6 lb/hr).

A 9.3.2 Special Consideration Affecting the Layout and Nature of the Heat Transfer Surface

The inlet and exit temperatures of the seeded combustion product flow bracket the fusion temperature of potassium sulfate [1342°K (1955°F)], which is the principal chemical species carrying the potassium seed at this point in the cycle. This means that unless special design measures are taken, the steam tubes, operating with a maximum external

wall temperature of 921°K (1200°F), will act as a cold trap for the potassium sulfate.

As potassium sulfate builds up on the tubes, the solid deposit layer provides a rising heat transfer impedance between the deposit-gas interface and the tube wall. Commensurate with this is a rising deposit-gas interface temperature. Providing the gas temperature is above the potassium sulfate fusion temperature, an equilibrium thickness of deposit would exist at which the deposit-gas interface temperature reached the fusion temperature. Ideally, then, further deposition would be in liquid form and could be drained off. If, however, the gas temperature were below the potassium sulfate fusion temperature, soft solid particles of potassium sulfate would be precipitated on the tube walls and in the gas stream. Those formed in the gas stream would eventually stick to the already encrusted tubes. Without the ability to drain off there would exist no equilibrium deposit thickness, resulting in severe impedance to flow and, in some areas, total blockage.

An attendant problem to that cited above is that liquid potassium sulfate is highly corrosive to any steels which might be considered for use in steam generator tubes. Hard solid particles would be a great deal less corrosive.

These considerations lead to the following three design axioms:

a. Wherever the seeded combustion products exist at temperatures above the potassium sulfate fusion temperature, the tubes should be coated with a high-temperature ceramic to protect them from severe corrosion.

b. Any bare tube surface should see only fully solidified particles of potassium sulfate. These could be removed from the tube surface, should they deposit, by conventional soot-blowing procedures.

c. There should be no region of the steam generator which is populated by steam tubes, ceramic coated or bare, wherein the gas temperature is at or very close to the potassium sulfate fusion temperature.

Axioms (b) and (c) together connote the existence of an open ductlike section of the steam generator in which the potassium sulfate is quenched into relatively hard solid particles by the addition of cool air. This quenching duct should be long enough, considering the gas velocity pertaining, to ensure complete solidification before the next tube section is encountered.

Experience in the Soviet Union and the United Kingdom has shown that when water at modestly critical pressures, 24.235 MPa (3515 psi) abs, flows downward in a tube and passes through the pseudocritical temperature [defined as T at which $(\frac{\partial\rho}{\partial T})_p$ is maximized], a severe temperature peaking of the tube wall might be experienced, providing the heat flux is high enough. This phenomenon is presumed to be caused by a radial distribution of density which gives rise to a bouyancy force distribution of the same magnitude as the shear force distribution. The net shear stress distribution, thus, might be drastically changed, with the effect that turbulent diffusivity, and consequently heat transfer coefficient, is drastically reduced. The problem does not exist in upflow or in horizontal flow because in these situations the bouyancy and shear force fields are not counteracting. A multistart tube bank, such as is conceived for use in the evaporator bank of the steam generator described here, would be composed mainly of horizontal tubing. There would be vertical sections, however, with ℓ/d of the order of 10. Conditions in these vertical sections are, in all likelihood, within the region where there exists a reasonable probability of tube wall temperature peaking in the downflow situation. The following fourth design axiom, therefore, is laid down.

d. In a section of the steam generator tube banks where the water passes through the pseudocritical temperature, the net flow vector should be upward not downward.

A 9.3.3 Temperature Approach Considerations

The relatively low combustion product exit temperature of 425°K (306°F) dictates that the feedwater should enter the steam generator where the gas exits.

One layout option had the evaporator section continuously running counterflow to the combustion products from the feedwater condition all the way to the exit superheat condition and the reheater bank being matched to the hottest gas conditions. Axiom (a), however, requires that until the gas temperature is reduced to some level just above the potassium sulfate fusion temperature, the tubes must be ceramic coated. The reheat load is not sufficient to bring the gas temperature down to this level, and we would not wish to quench the gas to below the potassium sulfate fusion temperature from a level higher than necessary; a portion of the main evaporator section, therefore, would be used in the prequench section and would require coating. There seemed to be a significant advantage from the aspect of simpler headering if only one tube bank were used in the prequench section. Accordingly, the reheat section is not used here; instead, about 25% of the main evaporator section is used to lower the combustion products from the inlet temperature of 1610°K (2438°F) to a prequench temperature of 1340°K (1952°F). This section is known as the finish superheat section. The reheater is then located postquench along with the first 75% of the main evaporator and is of bare tube construction.

It is important to note that the temperature of the main evaporating water stream as it crosses over from the postquenched gas stream to the prequenched gas stream is above the pseudocritical temperature. This means that, adhering to design axiom (d), water flow in the first 75% of the main evaporator (bare tube section) should be net upflow. The flow in the finish superheat section can be downward if required. Figure A 9.3.1 illustrates a steam generator layout which satisfies the constraints imposed by our four design axioms and temperature approach requirements. Figure A 9.3.2 is a temperature approach diagram.

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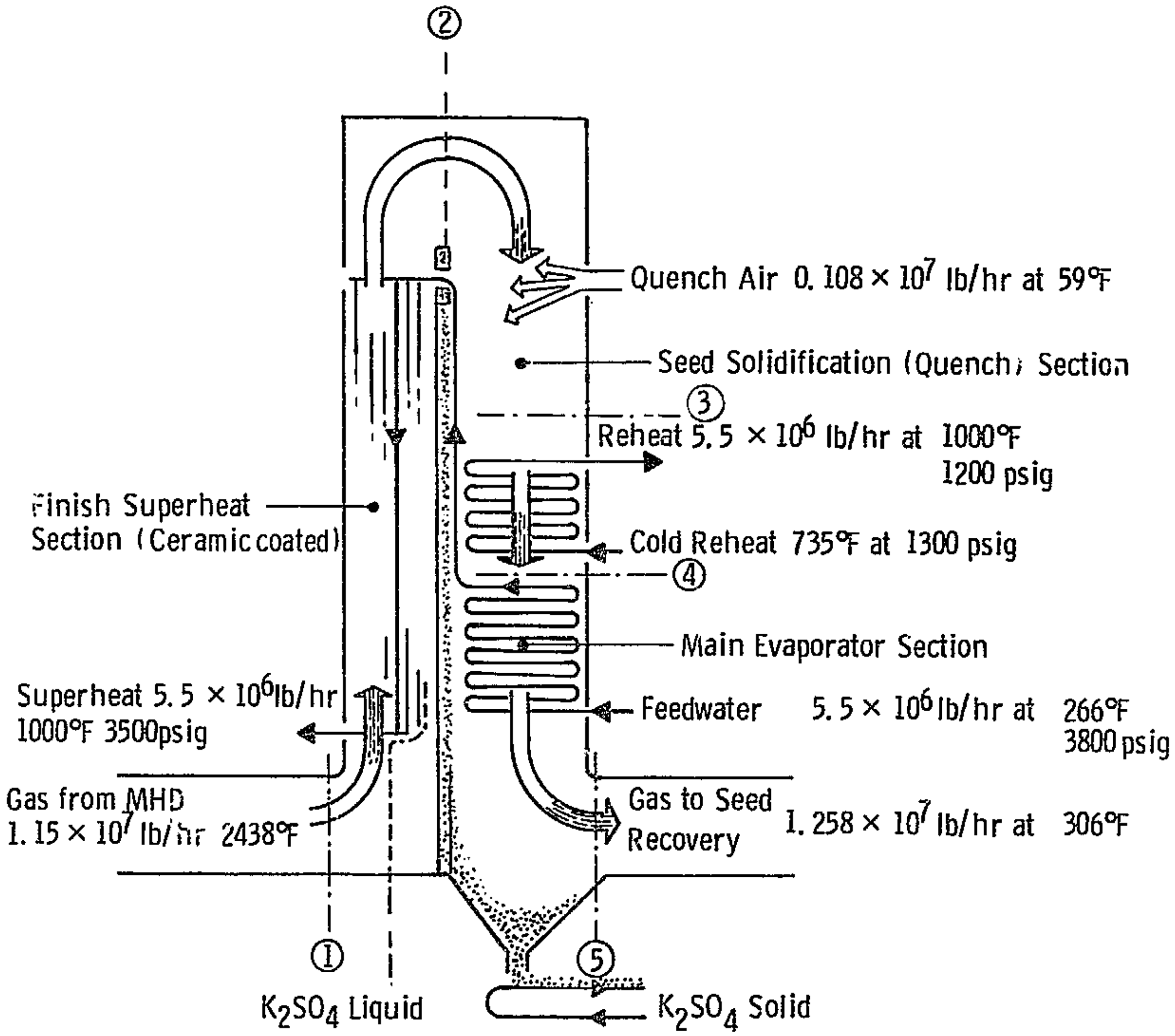

Fig. A 9.3.1—Steam generator schematic

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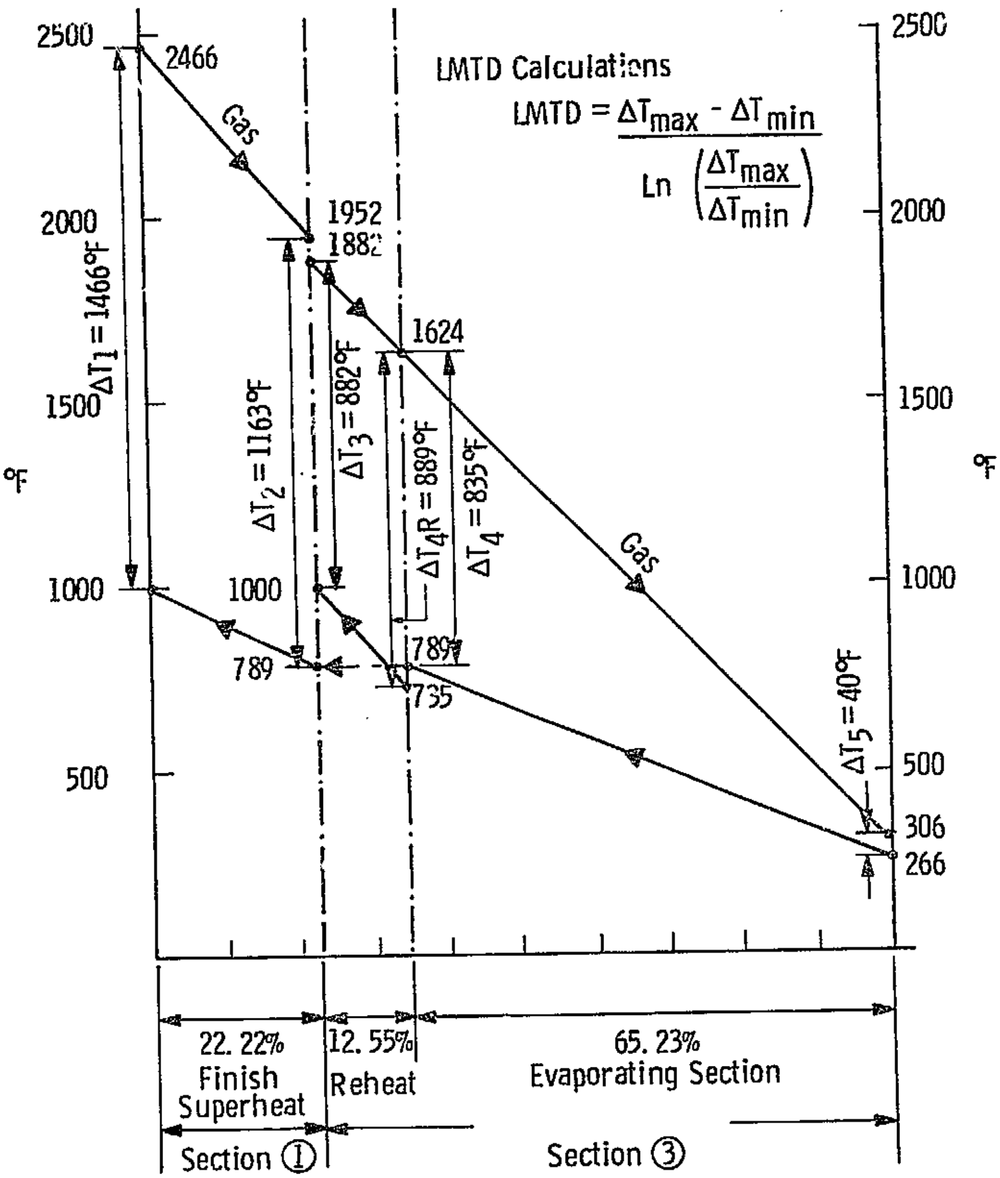

Fig. A 9.3.2—Steam generature temperature approach diagram

A 9.3.4 Description of the Finish Superheat Section and the Procedures Used to Determine the Mean Heat Transfer Coefficient

With the layout of heat transfer surface shown in Figure A 9.3.1 in mind, and considering that we wish to minimize header piping, it is advantageous to run the stream downward countercurrent with the combustion products in the finish superheat section.

Since this region exists before the quenching of potassium sulfate into solid form, it is clear that the liquid phase will impact and trap out on the heat transfer surface. This surface, therefore, must be a high-temperature ceramic such as a chrome-bonded alumina.

Preliminary estimates of the overall heat transfer situation indicated that the ceramic wall would exist at a temperature below the solidification temperature of potassium sulfate. Thus, a layer of solid potassium sulfate would build up on the ceramic and would reach an equilibrium thickness at which the interface with the gas stream reaches the fusion temperature. Further deposition of potassium sulfate would run off and could be collected. For this reason it was decided that the heat transfer surface should be in the form of vertically hung walls. Each wall would be essentially a slab of ceramic encasing a row of vertically hung steam tubes as per Figure A 9.3.3.

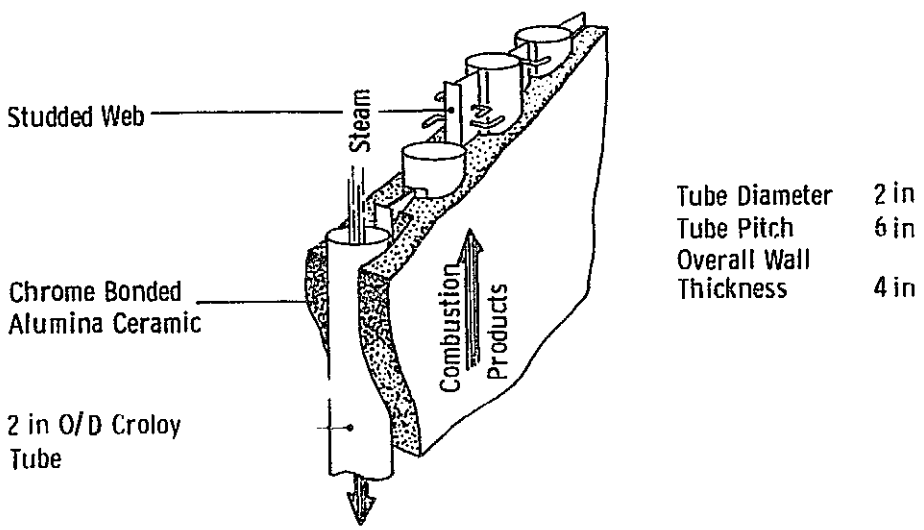

Figure A 9.3.3 - Section of finish superheat surface.

The determination of the heat transfer rate in the finish superheat section is simplified by the fact that one surface, namely the potassium sulfate slag-combustion product gas interface, is at a known-fixed temperature. This temperature, the potassium sulfate melting point, Tmp, is 1342°K (1955°F). We have then:

$$q,\ \mathrm{Btu/hr\text{-}ft^2} = h\ (\bar{T}_g - T_{mp}) \qquad \text{(A 9.3.1)}$$

In Equation A 9.3.1, h is the sum of radiative and convective components of heat transfer from the gas at its mean temperature over the section ($\bar{T}_g$) to the molten wall.

$$h_{fs},\ \mathrm{Btu/hr\text{-}ft^2\text{-}°F} = h_{rad} + h_{conv} \qquad \text{(A 9.3.2)}$$

We shall examine these components separately starting with h_{rad}.

The method used to evaluate h_{rad} is the classical one for radiation from a nonluminescent gas. This is presented in a number of tests including Siegal and Howell, and McAdams (Reference 9.25).

$$h_{rad},\ \frac{\mathrm{Btu}}{\mathrm{hr\text{-}ft^2\text{-}°F}} = \frac{\delta\left(\varepsilon_g(\bar{T}_g + 460)^4 - \alpha_{gmp}(T_{mp} + 460)^4\right)}{\bar{T}_g - T_{mp}} \qquad \text{(A 9.3.3)}$$

where ε_g is the effective emissivity of the gas at $\bar{T}_g$ and is the sum of the positive contributions of the water vapor, the carbon dioxide, and the carbon monoxide, and a negative contribution due to the overlap of the water vapor and carbon dioxide emission bands.

$$\varepsilon_g = (\varepsilon_{CO_2,\bar{T}_g})\ (C_{CO_2} + (\varepsilon_{H_2O,\bar{T}_g})\ (C_{H_2O}) + (\varepsilon_{CO,\bar{T}_g})\ (C_{CO}) - \Delta\varepsilon_{T_g} \qquad \text{(A 9.3.4)}$$

In Equation A 9.3.4, ε_{CO_2}, ε_{H_2O}, and ε_{CO} are determined by reading, for example, Figures 4.14, 4.15, and 4.22 of McAdams. These figures present ε as a function of T_g with the product of partial pressure and effective radiation path length as a parameter. The partial pressures are, of course, known and depend upon the choice of coal. For Base Case 1, the partial pressures were 15.7, 8.61, and 2.229 kPa (0.155, 0.085, and 0.022 atm) for carbon dioxide, water vapor, and carbon monoxide, respectively.

The radiation path length is given by:

$$L = \frac{(4)\ \text{(Cross-Sectional Area)}}{\text{Perimeter}} \qquad \text{(A 9.3.5)}$$

For a gas enclosed between infinite parallel plates (the situation in the finish superheat section) Equation A 9.3.5 becomes

$$L = \text{(2)(Plate Separation)} \qquad \text{(A 9.3.6)}$$

Following an iterative form of calculation, 1.219 m (4 ft) was selected as the wall separation; therefore,

$$L = 2.438 \text{ m (8 ft)} \qquad \text{(A 9.3.7)}$$

In Equation A 9.3.4, C_{CO_2}, C_{H_2O}, and C_{CO} are factors for use whenever the total pressure is substantially different from 101.3 kPa (1 atm) and, as such, are not used here.

$\Delta\varepsilon_{T_g}$ in Equation A 9.3.4, can be obtained from Figure 4.17 of McAdams (Reference 9.25). This figure plots $\Delta\varepsilon$ against $P_{H_2O}/(P_{H_2O} + P_{CO_2})$ with $L(P_{H_2O} + P_{CO_2})$ as a parameter.

The result of these calculations was to give, for Base Case 1,

$$\varepsilon_g = 0.275 \qquad \text{(A 9.3.8)}$$

Again, in the equation for radiation heat transfer coefficient (Equation A 9.3.3)

$$\alpha_{gmp} = \alpha_{CO_2,mp} + \alpha_{H_2O,mp} + \alpha_{CO,mp} - \Delta\alpha_{mp} \qquad \text{(A 9.3.9)}$$

where

$$\alpha_{CO_2,mp} = \varepsilon_{CO_2,T_{mp}} \left(\frac{\bar{T}_g + 460}{T_{mp} + 460}\right)^{0.65} \qquad \text{(A 9.3.10)}$$

and

$$\alpha_{H_2O,mp} = \varepsilon_{H_2O,T_{mp}} \left(\frac{\bar{T}_g + 460}{T_{mp} + 460}\right)^{0.65} \qquad \text{(A 9.3.11)}$$

and

$$\alpha_{CO,mp} = \varepsilon_{CO,T_{mp}} \left(\frac{\bar{T}_g + 460}{T_{mp} + 460}\right)^{0.65} \qquad \text{(A 9.3.12)}$$

In Equations A 9.3.10 through A 9.3.12, α is determined as described above, except that T_{mp} is used rather than $\bar{T}_g$ when using the figures. Likewise,

$$\Delta\alpha_{mp} = \Delta\varepsilon_{T_g} \qquad \text{(A 9.3.13)}$$

and this is found using Figure 4.17 of McAdams (Reference 9.25) as before.

The result of these calculations is to give, for Base Case 1,

$$\alpha_{g,mp} = 0.296 \qquad \text{(A 9.3.14)}$$

For Base Case 1, the arithmetic mean gas temperature over the finish superheat section, $\bar{T}_g$, is 1478°K (2200°F) and, as previously stated, the potassium sulfate fusion temperature is 1341°K (1955°F), (T_{mp}). Using these values in Equation A 9.3.3 along with those for ε_g and α_{gmp} stated above, we obtain

$$h_{rad} = 122.4 \text{ W/m}^2 \text{°K} \left[21.56 \frac{\text{Btu}}{\text{hr-ft}^2\text{-°F}}\right] \qquad \text{(A 9.3.15)}$$

In the convection heat transfer coefficient the correlation used is that of Dittus and Boelter (Reference 9.25).

$$h_{conv} = \frac{0.023 K_g}{d_e} \left(\frac{V \rho_g d_e}{\mu_g}\right)^{0.8} Pr_g^{0.4} \qquad \text{(A 9.3.16)}$$

In the above equation the combustion product gas properties are evaluated at the average bulk temperature for the section, namely, $\bar{T}_g$. The hydraulic equivalent diameter (d_e) is given by twice the wall separation, in other words, 2.438 m (8 ft).

Iterative calculations involving heat transfer and pressure drop and taking account of the desired general layout of the steam generator showed that the bulk gas velocity should be around 30.48 m/s (100 ft/s). With a total gas mass flow rate of 1449 kg/s (1.15×10^7 lb/hr) eight between-wall passages 1.219 m (4 ft) wide by 18.288 m (60 ft) deep provide sufficient flow area. The resulting convection heat transfer coefficient is:

$$h_{conv} = 20.89 \text{ W/m}^2\text{°K (3.68 Btu/hr-ft}^2\text{-°F)} \qquad \text{(A 9.3.17)}$$

Adding the radiation heat transfer coefficient, we obtain:

$$h = h_{rad} + h_{conv} = 143.29 \text{ W/m}^2{}^\circ\text{K} \ (25.24 \text{ Btu/hr-ft}^2\text{-}^\circ\text{F}) \quad \text{(A 9.3.18)}$$

The heat to be transferred in this section is;

$$\dot{Q}_{fs} = 516 \text{ MWt} \ (1.761 \times 10^9 \text{ Btu/hr}) \quad \text{(A 9.3.19)}$$

and the average temperature gradient across which heat is transferred is:

$$\Delta T_{fs} = \bar{T}_g - T_{mp} \quad \text{(A 9.3.20)}$$

$$\Delta T_{fs} = 2200 - 1955 = 245^\circ\text{F} \quad \text{(A 9.3.21)}$$

The wall surface area provided in the first superheat section is accordingly:

$$A_s = \frac{Q_{fs}}{\Delta T_{fs} \quad h_{fs}} \quad \text{(A 9.3.22)}$$

$$A_s = \frac{1.761 \times 10^9}{245 \quad 25.24} = 2.646 \times 10^4 \text{ m}^2 \ (2.848 \times 10^5 \text{ ft}^2) \quad \text{(A 9.3.23)}$$

Commensurate with eight interwall flow channels each 18.29 m (60 ft) deep, the height of the walls is given by Equations A 9.3.24 and A 9.3.25.

$$H = \frac{A_s}{(8)(2)(60)} \quad \text{(A 9.3.24)}$$

$$H = 90.4 \text{ m (297 ft)} \qquad \text{(A 9.3.25)}$$

It is not likely that two walls would be made up a single slab high; rather, it is likely that the finish superheat section would be made up of two or three sections, as per Figure A 9.3.4, with facilities for the collection of potassium sulfate at the bottom of each section.

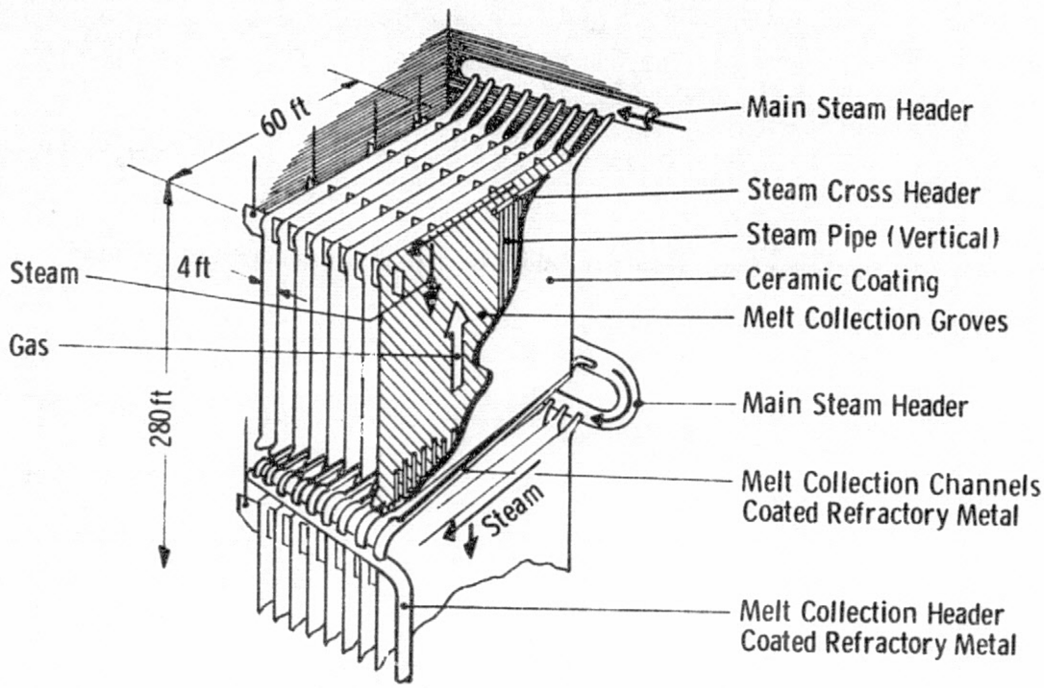

Figure A 9.3.4 Finish superheat section details.

A 9.3.5 Description of the Main Evaporator and Reheat Sections and Procedures Used to Determine the Mean Heat Transfer Coefficients

In both the main evaporator and reheat sections the water or steam (whichever is applicable) climbs through serpentine tube banks in net counterflow to the combustion product gas which, following solidification of the potassium sulfate by cold air injection, is moving vertically downward. See Figure A 9.3.1.

Most of the tube surface is in cross flow, and the situation is illustrated by Figure A 9.3.5.

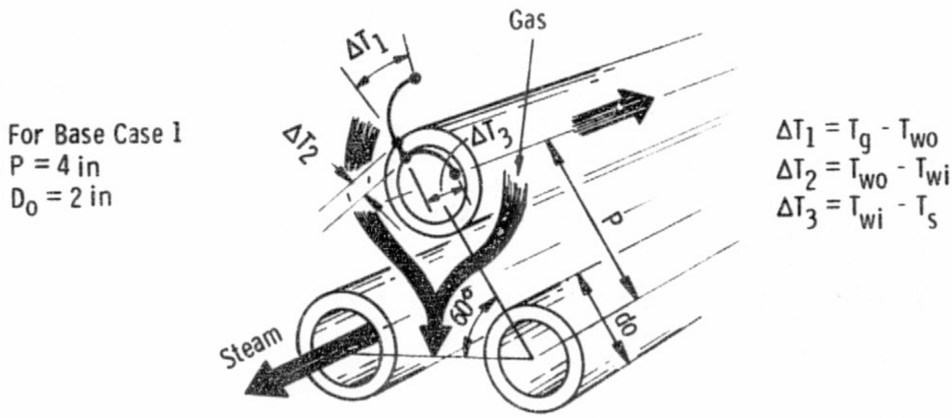

Figure A 9.3.5 - Detailed flow situation for evaporator and reheater.

The overall temperature difference between the gas and the steam is given by:

$$T_g - T_s = (T_g - T_{wo}) + (T_{wo} - T_{wi}) + (T_{wi} - T_s) \quad \text{(A 9.3.26)}$$

$$\frac{q}{h_o} = \frac{q}{h_g} + \frac{q}{h_w} + \frac{q}{h_s} \quad \text{(A 9.3.27)}$$

$$h_o = \frac{1}{\frac{1}{h_g} + \frac{d_o}{h_w d_m} + \frac{d_o}{h_s d_i}} \quad \text{(A 9.3.28)}$$

where d_o, d_i, and d_m are the tube wall outside, inside, and mean area per unit length of the tube.

For the heat transfer coefficient from the combustion product gas to the outer tube wall in cross flow, the following correlation is applicable:

$$h_g = 0.33 \left(\frac{k_g}{d_o}\right)\left(\frac{V_{max}\ \rho_g\ d_o}{\mu_g}\right)^{0.6} Pr^{0.3} \quad \text{(A 9.3.29)}$$

V_{max} is maximum velocity attained by the gas as it flows between the tubes. If the total mass flow rate of gas is $\dot{m}_g$ and each tube row contains N tubes of length ℓ feet, the expression for maximum velocity for a tube pitch, P, is

$$V_{max} = \frac{\dot{m}}{N(P - d_o)(\ell)(\rho_g)} \quad \text{(A 9.3.30)}$$

in both evaporator and reheater banks.

For Base Case 1, each tube row contains 180 tubes on 10.16 cm (4 in) pitch, 5.08 cm (2 in) diameter. The length of each tube is 19.81 m (65 ft). With a total gas mass flow rate of 1585 kg/s (1.258×10^7 lb/hr) the expression for V_{max} becomes:

$$V_{max} = \frac{1.796}{\rho_g} \text{ ft/s} \quad \text{(A 9.3.31)}$$

ρ_g is in lb/ft^3.

Table A 9.3.1 illustrates how one might proceed from Equation A 9.3.31 to deduce the required number of tube rows in the evaporator and reheat banks. This table also illustrates how the banks must be circuited in order to comply with steam-side pressure drop constants. Finally, it shows that the overall gas-side pressure drop would not exceed 10.13 kPa (0.1 atm). The other equation elements of Table A 9.3.1 are the steam-side heat transfer coefficient and the steam- and gas-side pressure drop.

For a complete review of the equations which have been proposed to deal with forced convective heat transfer to fluids above, but fairly close to, the critical pressure see Reference 9.36. Primarily because it gives the most conservative prediction, this writer prefers to use the correlation of Kutateladze and Leontiev which can be stated:

$$Nu_m = 0.023\, Re_m^{0.8}\, Pr_m^{0.4} \left[\frac{2}{\left(\frac{\rho_m}{\rho_w}\right)^{0.5} + 1}\right]^2 \qquad \text{(A 9.3.32)}$$

Subscript m indicates that the fluid properties are to be evaluated at a temperature which is the arithmetic mean of the bulk fluid and the wall temperatures. Subscript w indicates evaluation at the wall temperature.

For Base Case 1, the gas-side heat transfer is limiting, and the temperature difference between the bulk steam and the wall is just a few degrees. It is not necessary, therefore, to make any distinction between wall conditions and conditions at some average between the wall and the bulk. The Kutateladze-Leontiev equation (Equation A 9.3.32) then reduces to the more familiar Dittus-Boelter equation:

$$h_s,\ \text{Btu/hr-ft}^2\text{-°F} = 0.023 \left(\frac{k_s}{d_i}\right)\left(\frac{G\, d_i}{\mu_s}\right)^{0.8} Pr_s^{0.4} \qquad \text{(A 9.3.33)}$$

where G is the mass velocity.

$$G, \text{ lb/hr-ft}^2 = \frac{4\,\dot{m}_s}{\pi\, d_i^2\, N_{circ}} \qquad \text{(A 9.3.34)}$$

In Equation A 9.3.34 $\dot{m}_s$ is the total mass flow rate of steam or water in pounds per hour, and N_{circ} is the number of parallel circuits. The inside tube diameter d_i is in feet.

The equation used for the combustion product gas in cross flow over tubes is that recommended by the Heat Transfer Research Institute (HTRI) (Reference 9.37).

$$\Delta p, \text{ psi/tube row} = \frac{0.334\, f\, G_g^2}{10^{10}\, \rho_g} \qquad \text{(A 9.3.35)}$$

In Equation A 9.3.35, G_g is the gas mass velocity given by:

$$G_g, \text{ lb/hr-ft}^2 = \frac{12\,\dot{m}_g}{N(P - d_o)\,\ell} \qquad \text{(A 9.3.36)}$$

where $\dot{m}_g$ is the total gas mass flow in pounds per hour, N is the number of tubes per row, ℓ is the tube length in feet, and P and d_o are the tube pitch and outside diameter, respectively, in inches.

The friction factor, f, in Equation A 9.3.35 is a function of the tube arrangement, and Section C.2.1 of the HTRI design manual presents curves of f against Reynolds number for several tube patterns. For the tube arrangement selected for Base Case 1, the HTRI curve has been fitted, using Equation A 9.3.37.

$$f = 1.355805 - 1.068654 \log(Re) + 0.348888 [\log(Re)]^2$$

$$- 0.05109131 [\log(Re)]^3 + 0.002781413 [\log(Re)]^4 \quad \text{(A 9.3.37)}$$

In Equation A 9.3.37 Re is given by:

$$Re = \frac{G_g \ d_o}{12 \quad \mu_g} \quad \text{(A 9.3.38)}$$

For the flow of water or steam through smooth tubes, the equation used to determine the pressure drop for a length, ℓ, of pipe is based upon the following familiar expression which assumes a consistent set of units:

$$\frac{\Delta p}{\rho_s} = \frac{4f \ \ell \ V^2}{2g \ d_i} \quad \text{(A 9.3.39)}$$

where

$$V = \frac{4 \ \dot{m}_s}{\rho_s \ \pi \ d_i^2 \ N_{circ}} \quad \text{(A 9.3.40)}$$

Equation A 9.3.40 again assumes a consistent set of units.

When it is required to use the mass flow rate ($\dot{m}_s$) in lb/hr, the density (ρ_s) in lb/ft^3, the acceleration due to gravity (g) in ft/s^2, and the tube inside diameter (d_i) in inches, Equations A 9.3.39 and A 9.3.40 may be combined to give:

$$\Delta p = \frac{(64) \ 12^5 \ f \ \ell \ \dot{m}_s^2}{(144)(32.2)(3600^2)(\pi^2) \ d_i^5 \ \rho_s \ N_{circ}^2} \quad \text{(A 9.3.41)}$$

Depending upon the Reynolds number, f is obtained from one of the following equations which assume normal tube roughness:

If $Re \leq 10^4$

$$f = \frac{0.4517}{Re^{0.2939}} \qquad \text{(A 9.3.42)}$$

if $10^4 < Re \leq 2 \times 10^4$

$$f = \frac{0.3757}{Re^{0.2709}} \qquad \text{(A 9.3.43)}$$

if $2 \times 10^4 < Re \leq 5 \times 10^5$

$$f = \frac{0.02909}{(Re/10^4)^{0.179}} \qquad \text{(A 9.3.44)}$$

if $5 \times 10^5 < Re$

$$f = \frac{0.0159}{(Re/10^4)^{0.0245}} \qquad \text{(A 9.3.45)}$$

In Equations A 9.3.42 through A 9.3.45 Re is given by:

$$Re = \frac{G\ d_i}{\mu} \qquad \text{(A 9.3.46)}$$

or using $\dot{m}_s$ and the units as indicated previously

$$Re = \frac{48\ \dot{m}_s}{\pi\ d_i\ N_{circ}\ \mu_s} \qquad \text{(A 9.3.47)}$$

The equation for gas-side and steam-side heat transfer and pressure drop stated earlier form the basis of a computer program in which the heat transfer coefficient for the tube wall, referred to the outside diameter, is:

$$h_w = \frac{(12)(2)\ k_w}{\pi\, d_o\ \log\left(\frac{d_o}{d_i}\right)} \qquad \text{(A 9.3.48)}$$

Table A 9.3.1 indicates the route taken by the computer program in using the previous equations to define the main evaporator and reheat sections.

A 9.3.6 Costing of the Steam Generator for Base Case 1 and Method Used for Scaling Cost to Suit Other Design Conditions

This section is concerned with two topics:

- The determination of the capital cost of the steam generator for Base Case 1
- The formulation of a method for determining the capital cost of similar steam generators using Base Case 1 as a base.

For costing purposes the steam generator is referred to as Account 12. Ten major steam generator components have been identified and have been given subaccount numbers 12.1 through 12.10. The summation of these ten component costs forms the overall steam generator cost except that a 20% contingency is added to account for uncertainties. For purposes of reporting, these ten subaccounts were regrouped into three subaccounts.

The first, Subaccount 12.1, includes all items given in Category A in Table A 9.3.2 and in general refers to the superheater; the second, Subaccount 12.2, refers to the reheater and balance of boiler and

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TABLE A 9.3.1—DESIGN OF MAIN EVAPORATOR AND REHEAT TUBE BANKS USING 2 IN OD TUBES ON 4 IN EQUILATERAL PITCH

Reheater	Main Evaporator	Units	Section
1753	965	°F	Average Gas Temp, $\bar{T}_g$
0.0179	0.0283	lb/ft³	Gas Density at $\bar{T}_g$ ρ_g
100.3	63.34	ft/s	Max Gas Velocity $V_{max} = \frac{1.796}{\rho_g}$
19.12	16.49	Btu/hr ft² °F	Gas to Wall HTC, h_g See Eqn 9.3.29
0.00831	0.00515	psi/row	Gas Pressure Drop Per Row, ΔP/Row See Eqn 9.3.35
5.5×10^6	5.5×10^6	lb/hr	Total Steam Mass Flow, $\dot{m}_s$
180	180		Tubes Per Row, N_{Tube}
4 6	2 4 8		Number of Rows in Parallel, N_{Start}
720 1080	360 720 1440		Number of Parallel Circuits = $(N_{Tube})(N_{Start})$
539 389	2603 1495 858	Btu/hr ft² °F	Wall to Steam HTC h_s See Eqn 9.3.33
404 292	1952 1121 644	Btu/hr ft² °F	h Refered to OD $h_s' = \frac{d_i}{d_o} h_s$
10	12	Btu/hr ft² °F	Wall Thermal Conductivity k_w
418	501	Btu/hr ft² °F	Through Wall HTC $h_w = \frac{2\,k_w}{d_o \ln\left(\frac{d_o}{d_i}\right)}$
17.49 17.21	15.84 15.74 15.58	Btu/hr ft² °F	Overall HTC $h = \frac{1}{\frac{1}{h_g} + \frac{1}{h_w} + \frac{1}{h_s'}}$
885	262	°F	Log Mean Temp Dif $LMTD = \frac{(\Delta T_{max} - \Delta T_{min})}{\ln\left(\frac{\Delta T_{max}}{\Delta T_{min}}\right)}$

Reheater	Main Evaporator	Units	Section
0.9312×10^9	4.817×10^9	Btu/hr	Heat Transferred This Section, Q
6.016×10^4 6.114×10^4	1.161×10^6 1.168×10^6 1.180×10^6	ft²	Required Outside Tube Surface Area, $A_s = \frac{Q}{h\ LMTD}$
6126	6126	ft²	Area Provided Per Tube Row $A' = N_{tube}\,(\ell)\left(\frac{\pi d_o}{12}\right)$
9.8 9.98	189.5 190.7 192.6		Number of Tube Rows Required $N_{Row} = A_s/A'$
12 12	190 192 196		N_{Row} Upward Rounded to Make Evenly Divisible by N_{Start}
195 130	6175 3120 1592.5	ft	Length of Tube Per Circuit $\ell_{circ} = (N_{Row})(\ell)/N_{Start}$
0.84 0.3794	0.12123 0.03314 0.00938	psi/ft	Steam Pressure Drop Per Ft of Tube Length ΔP/ft See Eq. A 9.3.41
163.8 49.3	748.6 103.4 14.9	psi	Total Steam Pressure Drop $\Delta P_s = (\ell_{circ})(\Delta P/ft)$
← 6 →	← 4 →		Selected Design Indicated by ←No of Starts→
0.[illegible]	0.989	psi	Total Gas Pressure Drop $\Delta P_g = (\Delta P/Row)(N_{Row})$
1.089		psi	Combined Evaporator and Reheater Gas, ΔP
This Leaves Approximately 0.38 psi for Loss in Superheater Section in Cycles that $\frac{\Delta P}{P} \leq 0.1$			Comment
Rows 12 Tubes/Row 180 Starts 6 Tubes/Circuit 30 Row Length, ft 65	Rows 192 Tubes/Row 180 Starts 4 Tubes/Circuit 45 Row Length, ft 65		Summary of Tube Bank Size Parameters

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TABLE A 9.3.2—COST BREAKDOWN FOR OPEN-CYCLE MHD STREAM GENERATOR FOR BASE CASE 1

Sub account number	Catagory (See Note)	Description of Item	Weight, lb	$/lb Material	$/lb Labor	$/lb Total	Cost of Item $
11.1	A	Superheater tubes and headers (304 SS)	3×10^5	2.32	3.92	6.24	1,880,000
11.2	C	Reheater tubes and headers (T22)	7.5×10^5	1.62	0.60	2.22	1,665,000
11.3	B	Economizer and evaporator tubes and headers (T22)	12.5×10^6	1.62	0.87	2.49	31,120,000
11.4	C	Supplementary air injection					1,600,000
11.5	A	Superheat tube ceramic coating	7.1×10^5	0.32	0.15	0.5	355,000
11.6	C	Structural steel 11.6 a) Structure 11.6 b) Liner 11.6 c) Siding	14.7×10^6	0.28	0.1	0.38	5,580,000
11.7	C	Insulation	5×10^6	0.16	0.13	0.29	1,450,000
11.8	C	Soot Blowers					2,500,000
11.9	C	Ash Hopper					1,000,000
11.10	A	K_2SO_4 handling					2,000,000
Totals Less Contingency			33.86×10^6 lb				$ 49,150,000
Contingency 20%			6.77×10^6 lb (accounts for 11.4, 11.8, 11.9)				$ 9,830,000
Grand Totals			40.63×10^6 lb				$ 58,980,000

includes all items in Category C (This subaccount is mislabeled; it does not include the evaporator surface); the third, Subaccount 12.3, is mislabeled and includes all evaporator and economizer surface (Category B).

Table A 9.3.2 lists the subaccount components, along with their estimated weights where possible, and the estimated material and installation costs on a per pound or per foot of tube basis. Two items—the supplementary air injection blowers and the soot blowers—were costed on the basis of existing 1972 estimates for similar air moving equipment. These costs were prorated for capacity and were subjected to a 7.5% per year escalation for the period 1972 to 1974. For the ash hopper a lump figure was arrived at following consultations with the architect engineers, Chas. T. Main, Inc. For the potassium sulfate handling system a lump figure was used following consultations with Westinghouse engineers knowledgable in the field of seed removal and handling.

Category designations A, B, and C in Table A 9.3.2 are used in order that the overall steam generator cost for Base Case 1 can be broken down into three categories for purposes of extrapolating this base case cost to other designs.

The split between the heat output of the finish superheat section and the main evaporator section of the primary steam circuit is a function of the inlet gas temperature. Since the nature of the two sections is different, the cost of a particular case is dependent upon this split. The absolute cost of the primary steam circuit, as opposed to the division of cost between its components, is a function also of the mass flow rate of the gas.

The reheat steam circuit is a function only of the mass flow rate of the gas, since the seed solidification requirements dictate that the inlet gas temperature to this section is always the same.

Using inputed gas mass flow rates and temperatures pertinent to each case, the system computer program calculates the heat outputs of the various sections for the case in question. Suppose the superheater output of a particular case is determined and is called MW_{SH}. The superheat

output of Base Case 1 upon which all costs are based we know to be 516 MWt. The cost of the superheat section of the new case in question would then be given by Equation A 9.3.49.

$$\text{Cost of Superheater, \$} = (1.2)(\text{Superheater Cost Base Case 1})\left(\frac{MW_{SH}}{516}\right)^{0.88}$$

$$\text{Cost of Superheater, \$} = (1.2)(\text{Cost of Items Designated A})\left(\frac{MW_{SH}}{516}\right)^{0.88}$$

$$\text{Cost of Superheater, \$} = (1.2)(4.240 \times 10^6)\left(\frac{MW_{SH}}{516}\right)^{0.88} \qquad \text{(A 9.3.49)}$$

Similarly, if the output of the main evaporator section for a particular case is called MWe, and the output of the main evaporator section for Base Case 1 is known to be 1377 MWt, the Equation A 9.3.50 applies.

$$\text{Cost of Evaporator, \$} = (1.2)(\text{Evaporator Cost Base Case 1})\left(\frac{MWe}{1377}\right)^{0.88}$$

$$\text{Cost of Evaporator, \$} = (1.2)(\text{Cost of Items Designated B})\left(\frac{MWe}{1377}\right)^{0.88}$$

$$\text{Cost of Evaporator, \$} = (1.2)(31.15 \times 10^6)\left(\frac{MWe}{1377}\right)^{0.88} \qquad \text{(A 9.3.50)}$$

Likewise, if the total output of a particular case is called MWt and the output of Base Case 1 is known to be 2166 MWt, then

Equation A 9.3.51 applies. Note that the reheater section is included here since its output is tied only to the total output.

$$\text{Cost of Balance of SG Including Reheat, \$} = (1.2)\begin{pmatrix}\text{Cost of Balance of SG Including} \\ \text{Reheat Section for Base Case 1}\end{pmatrix}\left(\frac{MWt}{2166}\right)^{0.88}$$

$$\text{Cost of Balance of SG Including Reheat, \$} = (1.2)(\text{Cost of Items Designated C})\left(\frac{MWt}{2166}\right)^{0.88}$$

$$\text{Cost of Balance of SG Including Reheat, \$} = (1.2)(13.762 \times 10^6)\left(\frac{MWt}{2166}\right)^{0.88} \quad \text{(A 9.3.51)}$$

Combining Equations A 9.3.49, A 9.3.50, and A 9.3.51, we obtain an equation for the total cost of any particular case.

$$\text{Cost of Particular Parametric Point, \$} = (1.2)\left[4.240 \times 10^6 \left(\frac{MW_{SH}}{516}\right)^{0.88} + 31.15 \times 10^6 \left(\frac{MWt}{1377}\right)^{0.88} + 13.762 \times 10^6 \left(\frac{MWt}{2166}\right)^{0.88}\right] \quad \text{(A 9.3.52)}$$

For each particular case, or parametric point, the division of cost between material and labor, is assumed to be the same as for Base Case 1.

Appendix A 9.4

CYCLONE COMBUSTORS FOR MHD APPLICATION

A 9.4.1 Cyclone Combustor Design

The cyclone combustor was chosen as the best system for coal and as the system most easily adaptable to MHD. Although the cyclone furnace is generally run at atmospheric pressure, there are no obstacles to conversion for high pressures.

The combustor scheme has been patterned after those presently in use and based on the development work of British Coal Utilization Research Association (BCURA) and the U.S. Bureau of Mines (USBM). A sketch of the two-stage combustor is shown in Figure A 9.4.1. Preheated air will be supplied to the combustor and broken into three gas streams--the primary and secondary flow in the first stage, and the second-stage flow. The primary zone of the combustor will operate at an air equivalence ratio of 0.65.

Two methods of operating the multistaged cyclone furnaces to limit peak temperature have been considered. They are to use either a rich or a lean mixture in the first-stage cyclone. Although both methods limit the peak temperature in the first stage, operating the first stage fuel rich seems more advantageous. This method introduces all of the ash-bearing fuel in the low-temperature region, which promotes ash rejection and also simplifies the second stage, since only air must be introduced. The difficulties with this method are that the possibility exists of corrosive molten iron formation in the slag and of a combustion inefficiency due to free carbon formation.

The first method, running the first stage fuel rich, was selected because it offered the simplest system in design and operation. By operating fuel rich, all of the fuel can be injected into the first stage with only air

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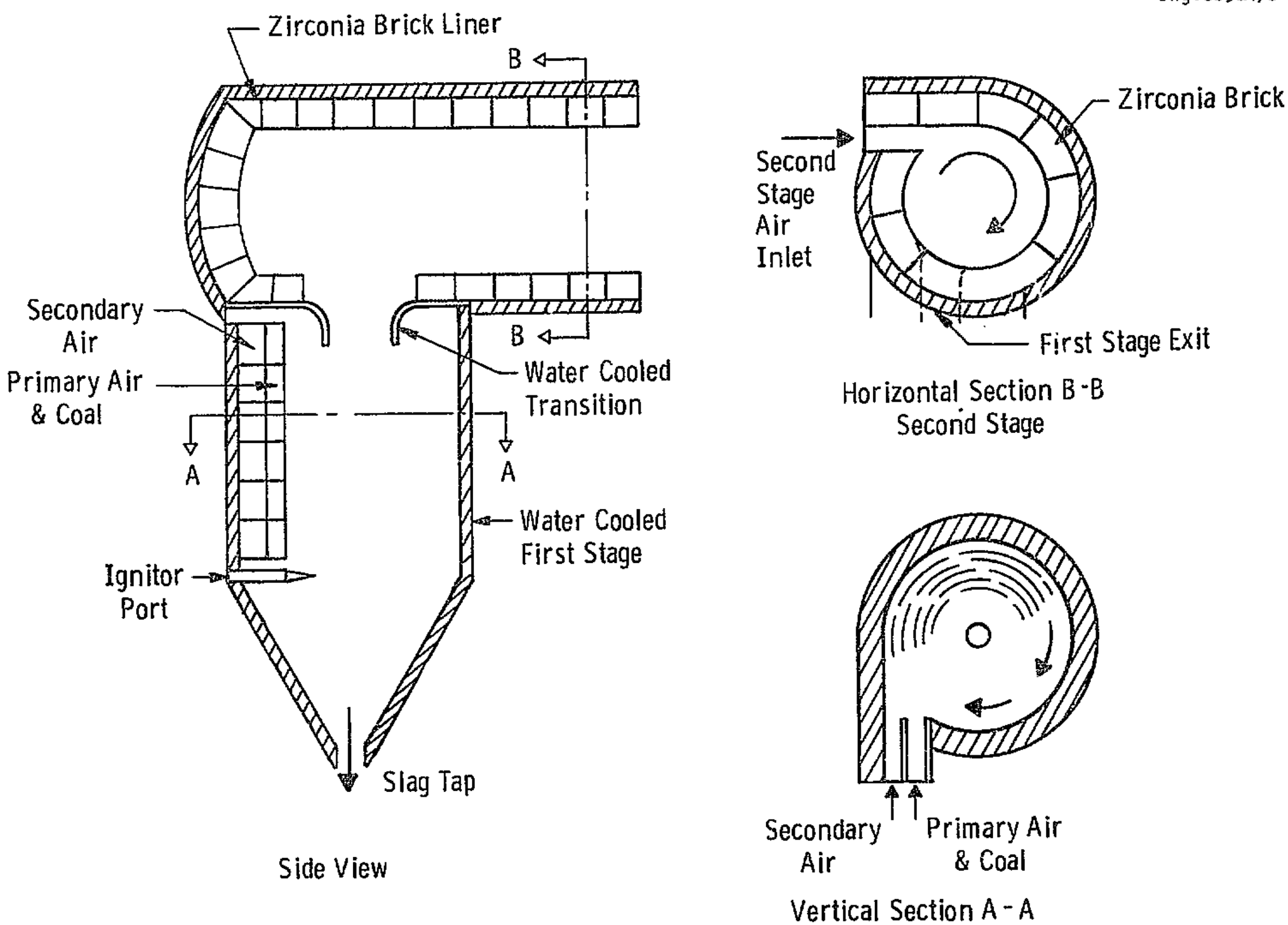

Fig. A 9.4.1—Two stage slagging combustor

being injected in downstream stages to reach the design operating temperature. This simplifies the process of fuel handling and distribution in the combustion systems. An air equivalence ratio of 0.65 was chosen because it is the lowest at which the iron in the slag remains oxidized as ferrous oxide (Fe_2O_3), and iron attack is eliminated. The air will enter the first stage in two streams designated primary and secondary airflows. The primary airflow will enter the combustor at an approximate velocity of 4.572 m/s (15 ft/s) along a secant (to avoid erosive action) and will act as a carrier for the pulverized fuel. The secondary airflow enters the combustor tangential to the inside wall at approximately 100 m/s (328 ft/s) and protects the wall from the erosive fuel particles while oxidizing the combustible portion of the fuel. When the air blanket has broken down, combustion is near completion and the ash portion of the fuel is in the liquid and vapor phase. Slag depostion on the water-cooled walls results in a frozen slag layer which provides the necessary thermal protection. The slag layer will reach an equilibrium thickness, and subsequent slag deposits will remain in the liquid vapor phase and run down the vertical combustor walls to the slag tap located at the bottom of the combustor cone. The temperature is maintained at the optimum value to afford a high enough slag viscosity [>0.25 Ns/m^2 (250 cp)] to afford sufficient runoff and yet vaporize as little slag as possible. Rejection rates as high as 93% are anticipated. At the combustor operating temperatures, all slag that is vaporized in the first stage is carried through the second stage, mixer, and duct in the vapor state. In the second stage, 30% of stoichiometric air is swirled tangentially to bring the air equivalence ratio to 0.95. In this manner a high fraction of the heating value of the fuel is utilized, but there is no free oxygen to form NO_x, which is a predominant pollutant at MHD operating temperatures. Ten percent more air is introduced to complete combustion and to utilize the entire heating value of the fuel at a point in the system where temperatures are not high enough to form objectionable amounts of NO_x.

The heat loss in the combustors will be taken as 5%. This heat is transferred to the steam plant feedwater so that some energy conversion occurs. Present literature indicates heat losses in the 5-to-10% range

based on atmospheric combustion. The heat relase in cyclone combustors is directly proportional to the pressure in atmospheres, and, therefore, an increase in pressure to MHD conditions will result in a substantial increase in heat release. The heat losses are not expected to increase with pressure, however, because the emissivity of the gases is close to unity at atmospheric conditions. Even considering the somewhat higher operating temperatures than those of conventional cyclone combustors, it was felt that an estimate of 5% heat loss was conservative.

The second stages and mixer are not water cooled, and no heat, therefore, is transferred to the steam plant. The heat loss in the second stage and mixer was chosen to be 0.2% because it was low enough not to affect the overall plant efficiency substantially and yet was relatively easy to obtain with a satisfactory amount of insulation. The required insulating brick thickness was determined for this heat flux by using manufacturers data on the thermal conductivity of brick presently available and by limiting the interface temperatures to levels that would prevent reactions. The brick chosen for the hot face was high-density zirconia because of its high operating temperature and excellent erosive characteristics. A more porous, less expensive zirconia brick is available, but it has poorer erosive characteristics. Magnesia brick is used behind the zirconia. The magnesia has a high thermal conductivity but provides the high-temperature, high-density requirements necessary in case of failure of the zirconia brick. The zirconia-magnesia interface is limited to 1811°K (2800°F). Several insulating bricks are available for the outside layer of insulation. Because of the high thermal conductivity of magnesium oxide, an insulating brick capable of operating up to 1811°K (2800°F) was chosen to minimize the required temperature drop across the magnesia. To obtain a standard shape and to provide a safe thickness of backup material for the zirconia, however, a standard magnesium oxide brick [24 by 5.08 by 22.85 cm (9-1/2 by 2-1/2 by 9 in)] was chosen. This results in the following brick thicknesses for a gas stream temperature of 2700°K (4400°F):

Dense zirconia	6.99 cm	2.75 in
Magnesia	6.35 cm	2.50 in
Insulating brick	4.45 cm	1.75 in

It was arbitarily decided that four combustor modules would be used. The combustors would be arranged so that they are joined on a mixer directly opposing another combustor. In this manner, the swirling flow used to increase combustion intensity in the combustors would be damped, and axial flow would be assured in the mixer and duct. The seed material would be injected in the mixer and would assume a uniform concentration before entering the MHD duct.

No combustion system (whether it be one-, two-, or three-stage) will be lined with a ceramic material when it is operating on a solid fuel containing ash because of the unavailability of a material that can withstand the severe corrosion problems encountered with the ash. The metal combustor walls will rely on the frozen slag layer for protection from the corrosive atmosphere and slag. When the combustor is operating on a clean fuel (coal gas), however, the wall will be protected by a silicon carbide liner.

Although a slagging-type combustor is not needed for the Base Case 3 points, a vortex-type combustor design will still be used because of its advantages of simplicity of design and ease of operation at high temperatures. The design can be simplified to have the following configuration for gas operation .

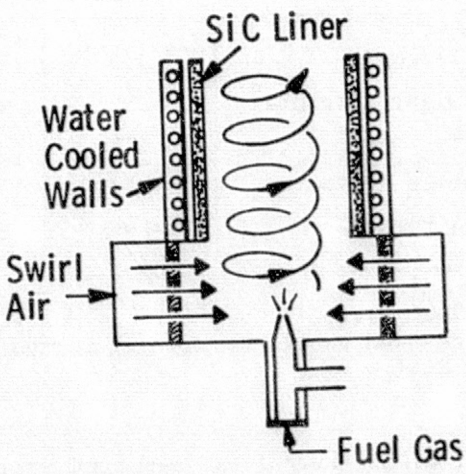

Figure A 9.4.2 - Base Case 3 single-stage combustor section

Here, a swirling combustion airflow will protect the ceramic wall from direct contact with the hot combustion products and, thus, allow a less costly refractory for insulation (silicon carbide). The combustors will be opposing. mounted horizontally, and fired into the mixer in the same manner as when operating on coal. The only significant design differences between operating on coal or char and on gas is the absence of ash, which is corrosive and erosive.

A 9.4.2 Combustor and Mixer Costing

Costs were found for the combustors and mixers using manufacturers data and installation costs from the A/E. A multiplier of 0.85 w s used in costing the zirconia brick because a 15% discount was to be applied to large orders. Silicon carbide was used in Base Case 3 because of the absence of corrosive slag. The densities and costs of the ceramics used are shown in Table A 9.4.1.

Table A 9.4.1 - Densities and Costs of Ceramics

	Density (lb/ft^3)	Cost ($\$/ft^3$)
Dense Zirconia	250	826.20
Lightweight Zirconia[a]	155	642.60
Magnesia	179	23.38
Insulating Brick	48	7.65
Silicon Carbide	195	273.00

[a]Lightweight zirconia was included for future reference even though it was not included in the present design.

The installation cost of the bricks was supplied by the A/E as \$331/Mg (\$150/1000 lb) regardless of brick size and weight.

Steel requirements were found by using wall thickness of 3.17 cm (1.25 in) and a density of 7865 kg/m^3 (491 lb/ft^3). Structural steel requirements are 0.30 kg of structural steel/kg of load. Installation cost of steel was $331/Mg ($150/1000 lb).

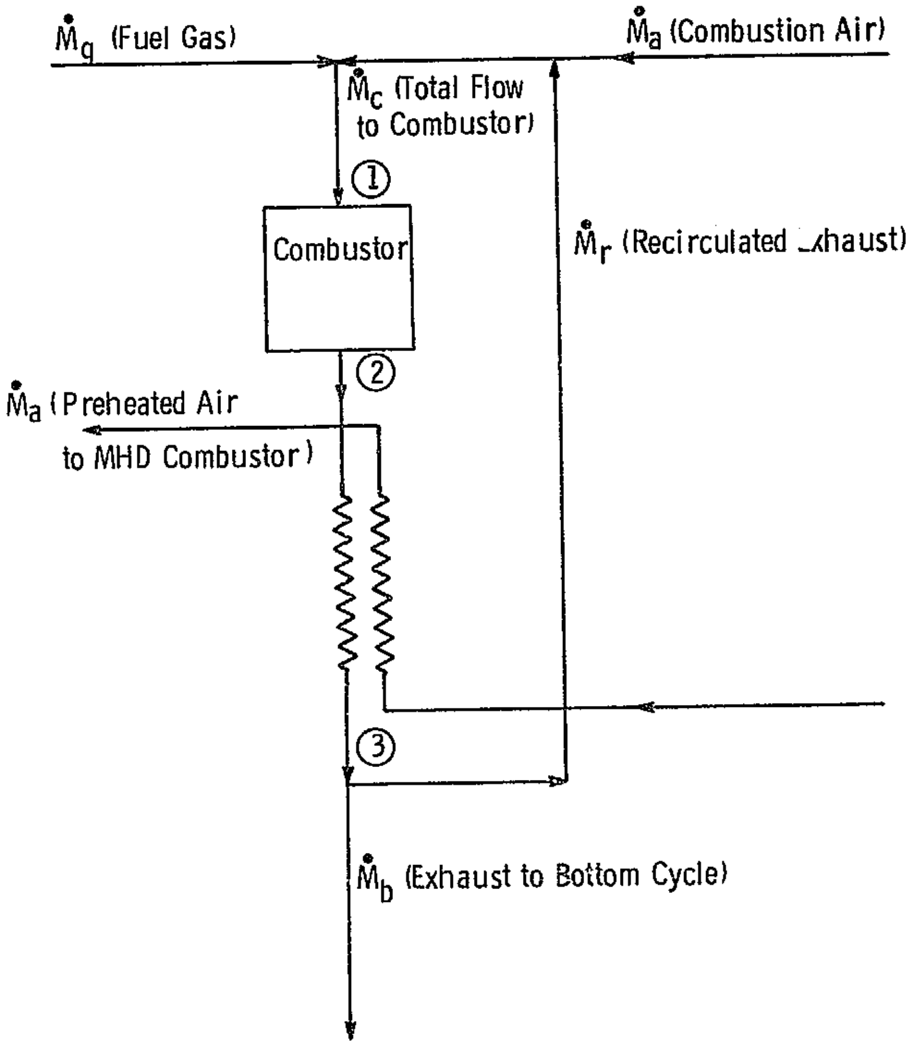

Fig. A 9.5.1—Schematic of separately fired air heater flow streams

Appendix A 9.5

A HIGH-TEMPERATURE HIGH-FLOW JET PUMP FOR APPLICATION TO A MHD TOPPING CYCLE

A 9.5.1 Introduction

The combustion products from the separately fired air heater are mixed with the exhaust from the MHD duct during its passage through the bottoming cycle heat exchangers. To avoid excessively high combustion temperatures, however, and associated design and maintenance problems, a portion of this exhuast is recirculated to the inlet of the air heater combustion chamber. In the schematic diagram (Figure A 9.5.1), it is the temperature at Point 2 that is of interest. As seen from the energy and materials balance, the enthalpy at this point is:

$$H_2 = \frac{\dot{M}_g H_g + \dot{M}_a H_a + \dot{M}_r H_r + \Delta H_c \dot{M}_g}{\dot{M}_a + \dot{M}_g + \dot{M}_r} \qquad \text{(A 9.5.1)}$$

where H refers to enthalpy and $\dot{M}$ refers to mass flow rate, with subscripts g, a, r denoting the gas, air, and recirculation streams, respectively. The quantity ΔH_c is the heat combustion of the gas. Thus, the temperature of the combustion products entering the heat exchanger section will decrease with increasing recirculation rates. A maximum temperature of 2255°K (3600°F) is indicated by the properties of economically available ceramic materials for construction. Recirculation rates up to 1.87 times the inlet airflow rate may be required to satisfy this limitation.

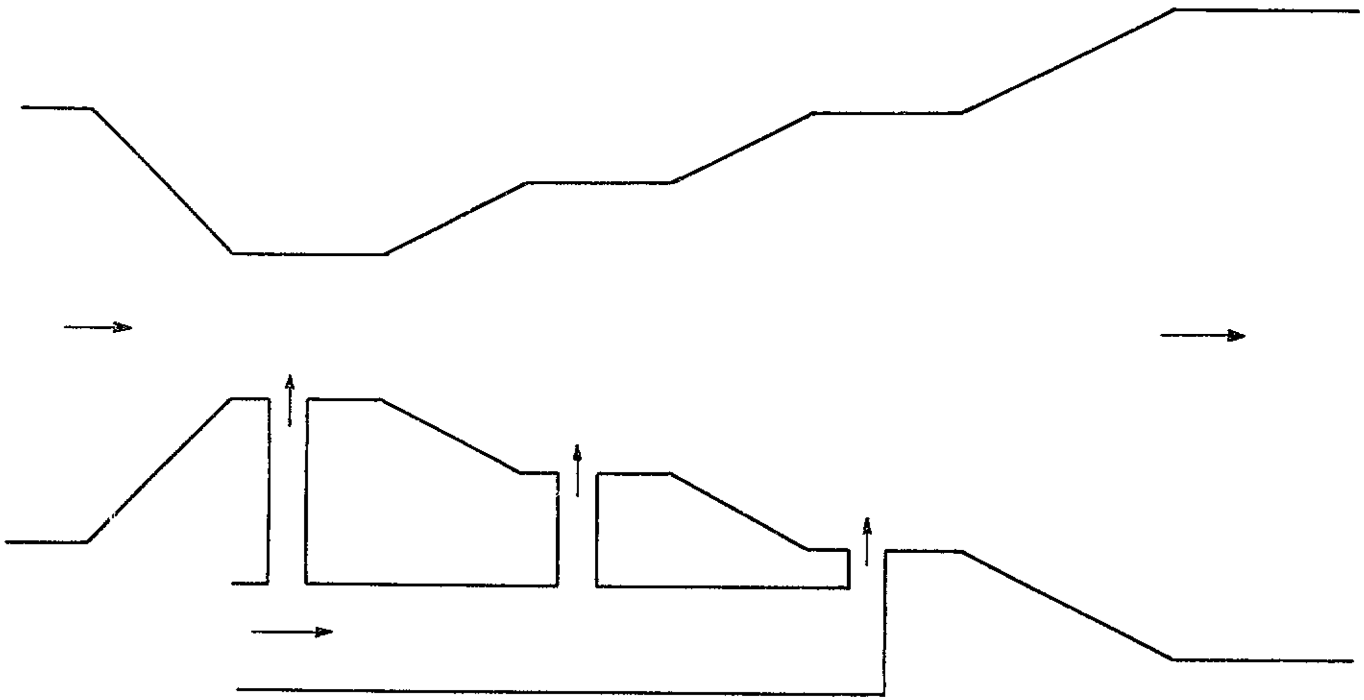

Fig. A 9.5.2—Diagram of multistage constant area jet pump

A 9.5.1.1 Basic Principles

In order to recirculate a portion of the flue gases from the separately-fired air heater back into its combustion chamber, this stream must be compressed back to inlet pressure. This means that the compressor must overcome the pressure losses in the chamber and in the ducting. This exhaust gas is nominally near atmospheric pressure but at a temperature of about 1680°K (2564°F), which precludes use of any conventional mechanical blower. It is believed that use of the jet pump principle will probably be required, using the inlet fresh air stream to induct the flue gas stream. A jet pump, however, has serious limitations in efficiency and capacity.

In concept the mixing of the two streams in a jet compressor may take place either in a passage of constant cross section or at constant pressure in an isobaric chamber. The first type of compressor is little used in practice but has been considered here because of additional structural and arrangement options. In particular, this concept avoids the necessity of an inlet stream nozzle, with substantial pressure differential across walls exposed to high-temperature gas streams on both sides. Such a nozzle would have to be constructued of refractory brick without steel reinforcement.

A 9.5.2 Constant Area Mixing

Staging is required in this type of jet compressor, since the ratio of outlet to inlet flow rates for any one stage may not exceed a certain value. Ideally, this ratio is $\sqrt{2}$ but is reduced by parisitic pressure losses in the system. For this application it is concluded that many stages would be required.

A staged jet pump is shown schematically in Figure A 9.5.2, where each stage consists of a throat at intake pressure containing an induction port, followed by a diffuser section for expansion back to

intake pressure, then followed by the throat of the next section. This assembly of stages is preceeded by an inlet nozzle and followed by the final diffuser. Based on the simplified assumptions of incompressible flow and constant fluid density, it may be shown that the overall static pressure drop across the pump assembly is:

$$P_0-P_2 = P_1-P_2-\frac{1}{2C_n}\frac{W_0^2}{\rho}+\frac{\frac{1}{2C_n}\left[\frac{W_0^2}{\rho}+\frac{2}{C_d}(P_2-P_1)\right]}{\prod_{n=1}^{N}\left[\frac{2}{C_d}+\alpha_n^2\left(1-\frac{2}{C_d}\right)\right]} \qquad (A\ 9.5.2)$$

where P = static pressure

ρ = density

W = mass velocity

C_d = diffuser efficiency

C_n = nozzle efficiency

α_n = inlet flow/outlet flow for the nth stage

N = number of stages

0 - refers to inlet to the nozzle

1 - refers to induction stream

2 - refers to outlet stream

Continuity requires that

$$\frac{M_2}{M_1} = \prod_{n=1}^{N} \alpha_n \qquad \text{(A 9.5.3)}$$

The overall pressure loss is minimized by setting

$$\alpha_n = \alpha = \text{constant.}$$

Thus $\alpha = \left(\frac{M_2}{M_1}\right)^{\frac{1}{N}}$ and

$$P_0 - P_2 = P_1 - P_2 - \frac{1}{2C_n}\frac{W_0^2}{\rho} + \frac{\frac{1}{2C_n}\left[\frac{W_2^2}{\rho} + \frac{2}{C_d}(P_2 - P_1)\right]}{\left[\frac{2}{C_d} + \alpha^2\left(1 - \frac{2}{C_d}\right)\right]^N} \qquad \text{(A 9.5.4)}$$

Since the quantity

$$\frac{2}{C_d} + \alpha^2\left(1 - \frac{2}{C_d}\right) \qquad \text{(A 9.5.5)}$$

must always be positive, the minimum number of stages is

$$N = \frac{2\,\ln\left(\frac{M_2}{M_1}\right)}{\ln\left(\frac{2}{2 - C_d}\right)} \qquad \text{(A 9.5.6)}$$

Say for $\frac{M_2}{M_1} = 2$, $C_d = 0.85$, $C_n = 0.95$.

$$N_{min.} = 2.505$$

that is, at least three stages would be required in order to achieve induction at all. Further examination shows that at such high induction rates a much larger number of stages is required for reasonable pressure losses.

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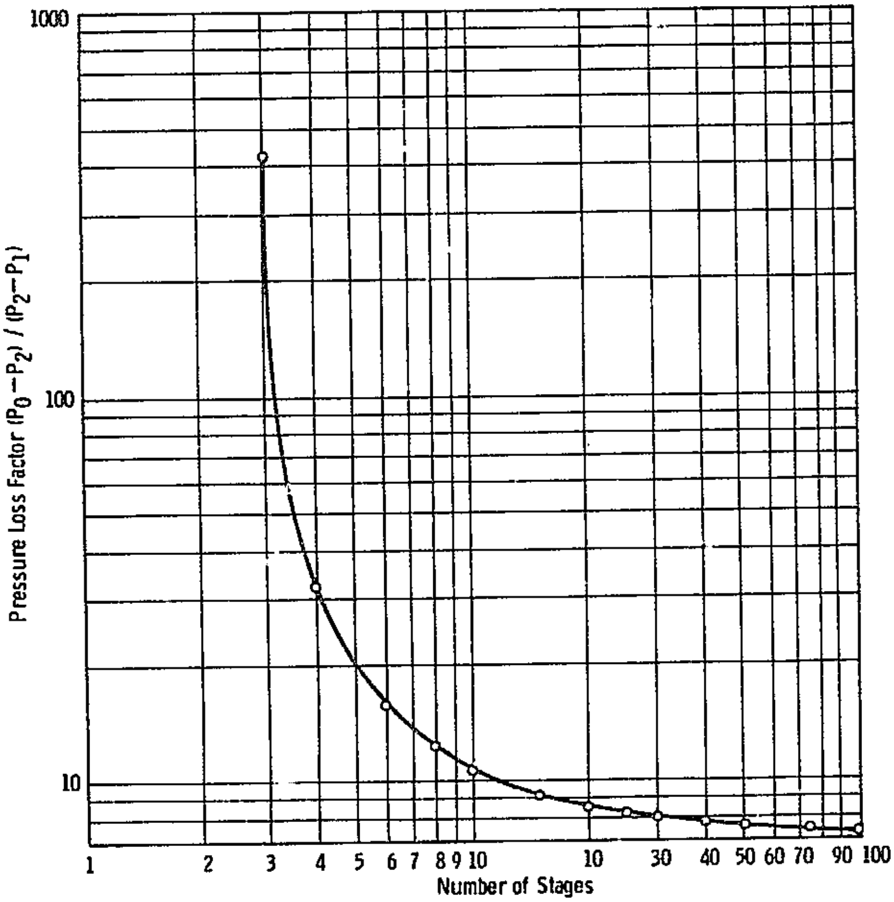


Fig. A 9.5.3—Pressure loss factor $(P_0 - P_2)/(P_2 - P_1)$ versus number of jet pump stages for flow ratio of 2.0

Figure A 9.5.3 shows the variation of the pressure loss factor $(P_0-P_2)/(P_2-P_1)$, neglecting inlet and exit velocity heads, as a function of number of stages, with $M_2/M_1 = 2$. The curve in this figure suggests that constant area mixing is not practical for the pumped flow rates required in this application.

A 9.5.3 Constant Pressure Mixing

The following relations describe the performance of a jet compressor with constant pressure mixing. These equations may be derived by direct solution of the energy and momentum equations, but it should be noted that they may also be obtained by taking the limit, as the number of stages approaches infinity, of the staged jet pump with constant area mixing.

$$\frac{W_1^2}{\rho} = \left(\frac{M_2}{M_1}\right)^{\frac{4}{C_d} - 2} \left[\frac{W_2^2}{\rho} + \frac{2}{C_d}(P_2-P_1)\right] \qquad \text{(A 9.5.7)}$$

$$\frac{W_3^2}{\rho} = \left[\frac{W_2^2}{\rho} + \frac{2}{C_d}(P_2-P_1)\right] \qquad \text{(A 9.5.8)}$$

$$P_0-P_2 = \frac{1}{2C_n}\left[\left(\frac{M_2}{M_1}\right)^{\frac{4}{C_d} - 2}\left\{\frac{W_2^2}{\rho} + \frac{2}{C_d}(P_2-P_1)\right\} - \frac{W_0^2}{\rho}\right] - (P_2-P_1). \qquad \text{(A 9.5.9)}$$

Here the subscript 3 refers to the inlet of the final diffuser.

The overall efficiency of the inductor system including the contribution of the compressor, η_c, is given by

$$\eta_P = \frac{\left(\frac{M_2}{M_1} - 1\right)(P_2 - P_1)\eta_c}{\frac{1}{2C_n}\left[\left(\frac{M_2}{M_1}\right)^{\left(\frac{4}{C_d} - 2\right)}\left\{\frac{W_2^2}{\rho} + \frac{2}{C_d}(P_2 - P_1)\right\} - \frac{W_0^2}{\rho} - (P_2 - P_1)\right]} \qquad \text{(A 9.5.10)}$$

It is expected that inlet and exit velocity heads will be small, and can be neglected, thus

$$\eta_p = \frac{\left(\frac{M_2}{M_1} - 1\right)\eta_c}{\frac{1}{C_d C_n}\left(\frac{M_2}{M_1}\right)^{\left(\frac{4}{C_d} - 2\right)} - 1} \qquad \text{(A 9.5.11)}$$

With that assumption and letting

$\eta_c = 0.9$

$C_n = 0.95$

$C_d = 0.85$

Figure A 9.5.3 shows the variation of pump efficiency, η_p, and pressure loss factor, $(P_0-P_2)/(P_2-P_1)$ as a function of flow ratio, M_2/M_1.

The estimated pressure loss in the combustion chamber, stone matrix, and ducting is 2%, to which an additional 0.4% has been added for flow control by throttling. Thus $P_2 = P_1 \times 1.024$.

Using this value we now calculate the inlet nozzle pressure ratio, P_1/P_0 which is also shown in Figure A 9.5.4.

The recirculation ratios used in the base case analyses result in the values of $M_2/M_1 = 1.70$, 2.02, and 2.87.

To obtain reasonably good pump efficiency, it is necessary to shape the isobaric chamber for efficient simultaneous mixing and pressure recovery. This, in practice, is largely empirical. It is anticipated that shapes practically attainable with firebrick construction may severely limit pump performance.

For present purposes of cost and performance estimates we have chosen a design which is structurally a compromise between staged constant-area mixing and constant-pressure mixing. This concept, as shown in Figure A 9.5.5 substitutes a conical diffuser with a multiplicity of induction ports for the isobaric chamber.

It is possible that problems may arise with the highest flow ratio (2.87) considered since the nozzle pressure ratio (as calculated on an incompressible basis) is approaching the critical pressure ratio 0.53, and the value calculated depends on maintaining good diffuser efficiencies in the presence of high induction rates, high Mach numbers, and the geometry limitations imposed by brick construction. (See Figure A 9.5.6 for the variation of P_1/P_0 as a function of C_d). The problems associated with restricting M_2/M_1 to 2.02 for cases 1, 2, 3, 15, and 16 are considered.

A 9.5.4 Design Parameters

It is assumed in the estimates that follow that the recirculation flow and the fresh airflow into the separately fired heater are each broken into two streams with two separate inductors.

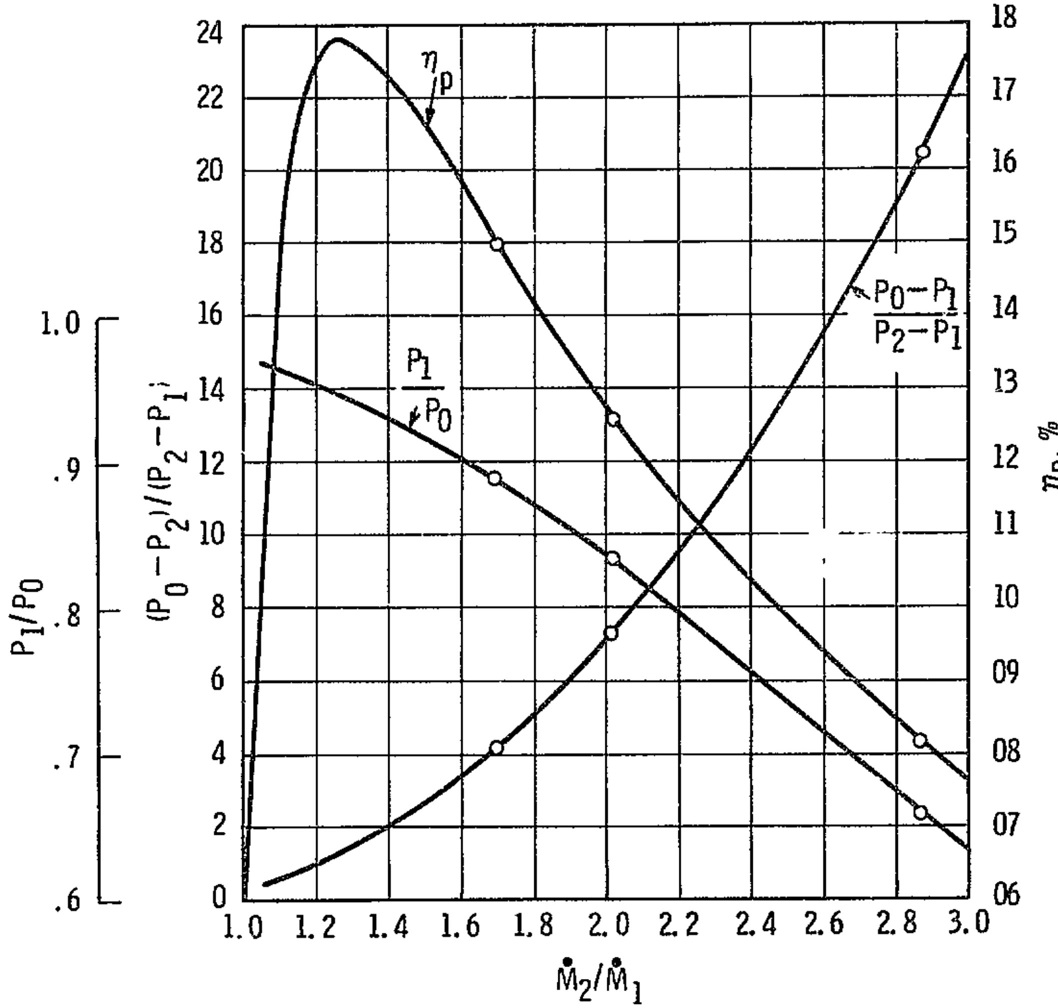


Fig. A 9.5.4—Nozzle pressure ratio pump pressure loss factor and pump efficiency versus flow ratio

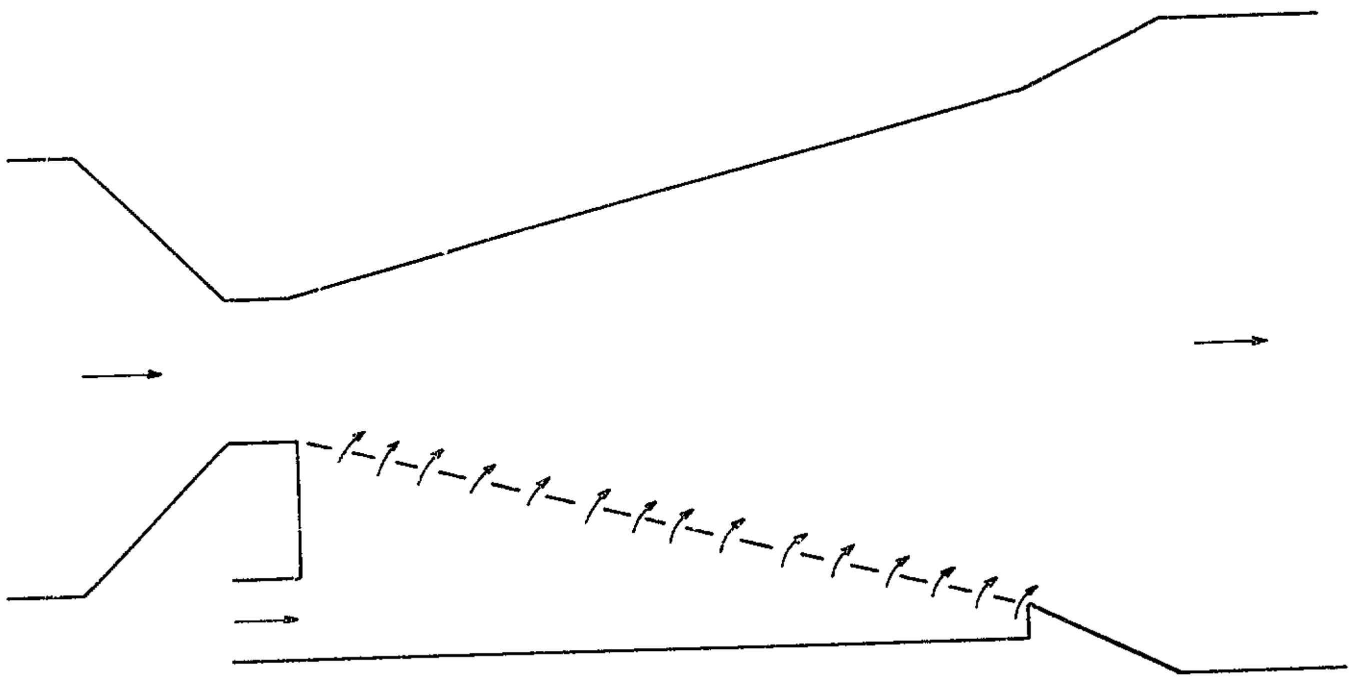


Fig. A 9.5.5—Continuously staged jet pump

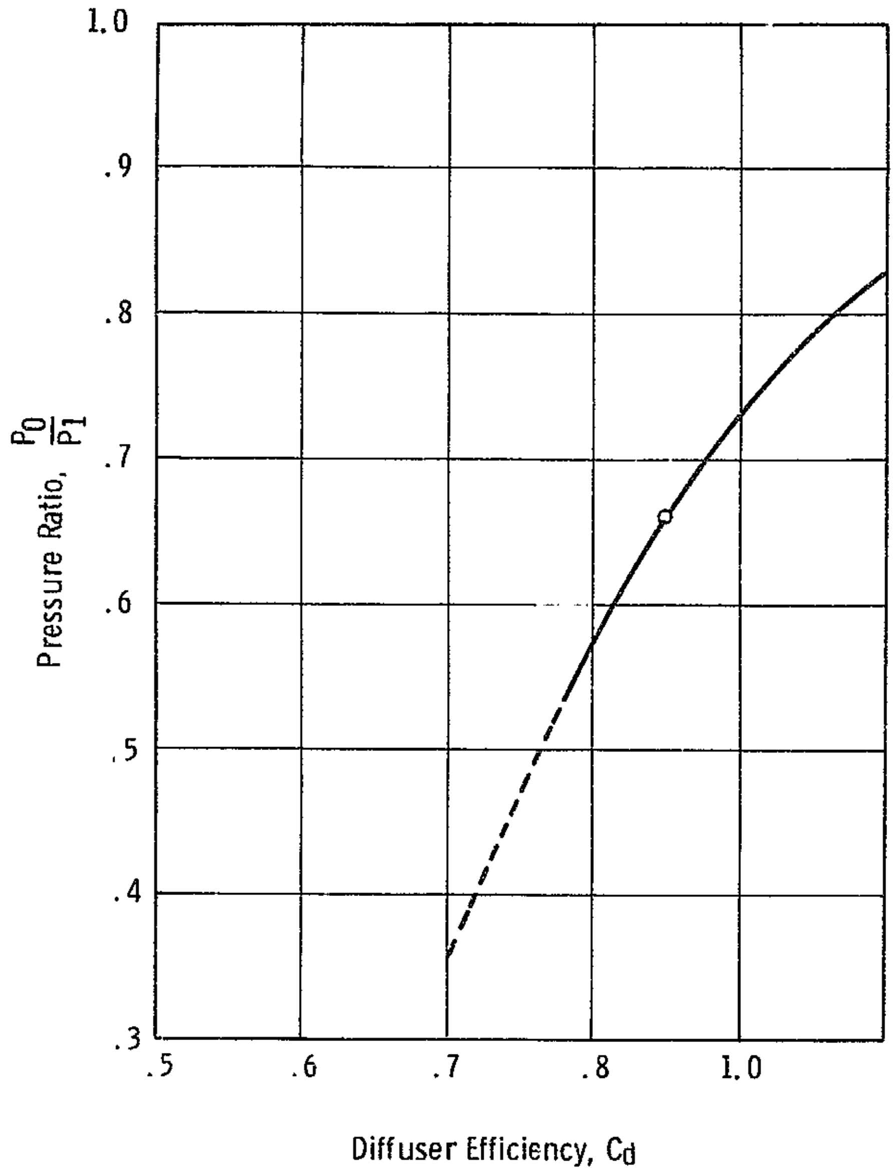


Fig. A 9.5.6—Nozzle pressure ratio versus diffuser efficiency for flow ratio 2.87 to 1

The basic design parameters are given in Table A 9.5.1 for each of those base cases that are unique in terms of the MHD topping cycle and that employ exhaust gas recirculation. In this table the nozzle inlet area and the diffuser exit area are based on limiting the velocity heads in these regions to 304 Pa (0.003 atm) or about 12.5% of the pressure difference that the pump must overcome.

Table A 9.5.1 Design Parameters

Case	M_1, lb/s	$\frac{M_2}{M_1}$	Nozzle Inlet Area, ft^2	Throat Area, ft^2	Area at End of Induction Section, ft^2	Diffuser Exit Area, ft^2
1	314.0	2.87	107.2	9.24	113.7	339.9
2	189.6	2.87	64.9	5.78	66.9	205.2
3	96.05	2.87	32.9	2.82	34.1	104.0
12	311.0	2.02	106.3	14.60	77.3	236.2
13	330.5	1.70	119.9	19.81	69.1	212.4
15	314.5	2.87	107.8	9.28	111.0	340.4
16	314.5	2.37	107.8	9.25	111.0	340.4
1	314.0	2.02	107.2	14.72	80.0	239.2
2	189.6	2.02	64.9	9.22	47.1	144.4
3	46.05	2.02	32.9	4.51	24.0	73.2
15	314.5	2.02	107.8	14.78	78.1	239.6
16	314.5	2.02	107.8	14.78	78.1	239.6

Base Case 1 with flow ratio 2.02 has been sized in more detail for purposes of cost estimating. The nozzle, the inductor-diffuser region, and the final diffuser region are all assumed to be conical sections of the cross section shown in Figure A 9.5.7.

This shape should be amenable to brick construction and offers only slightly more flow resistance than a circular cross section. A conceptual diagram of the brickwork is shown in Figure A 9.5.8. The

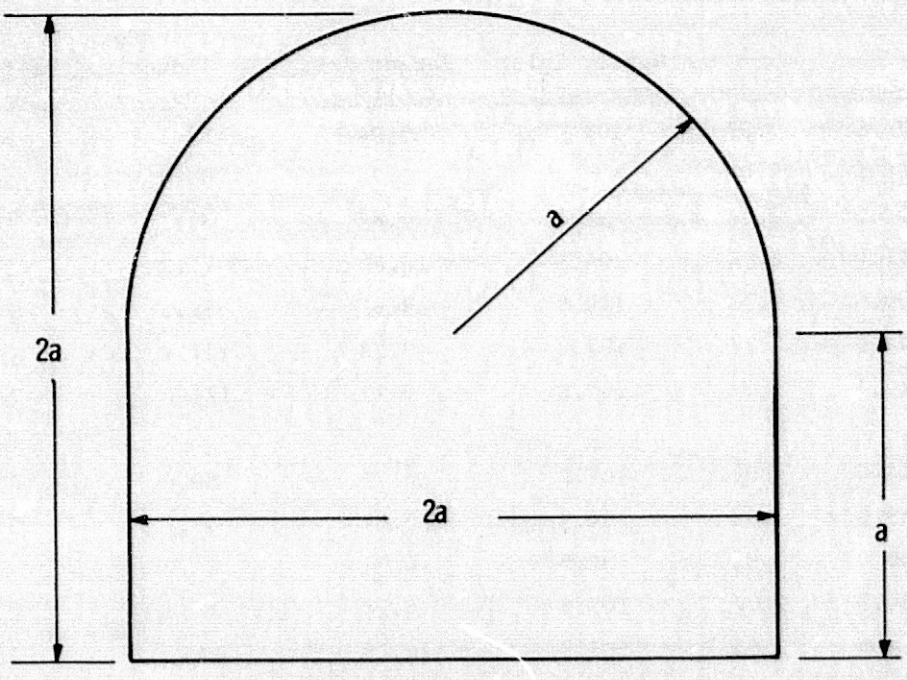


Fig. A 9.5.7— Jet pump cross section

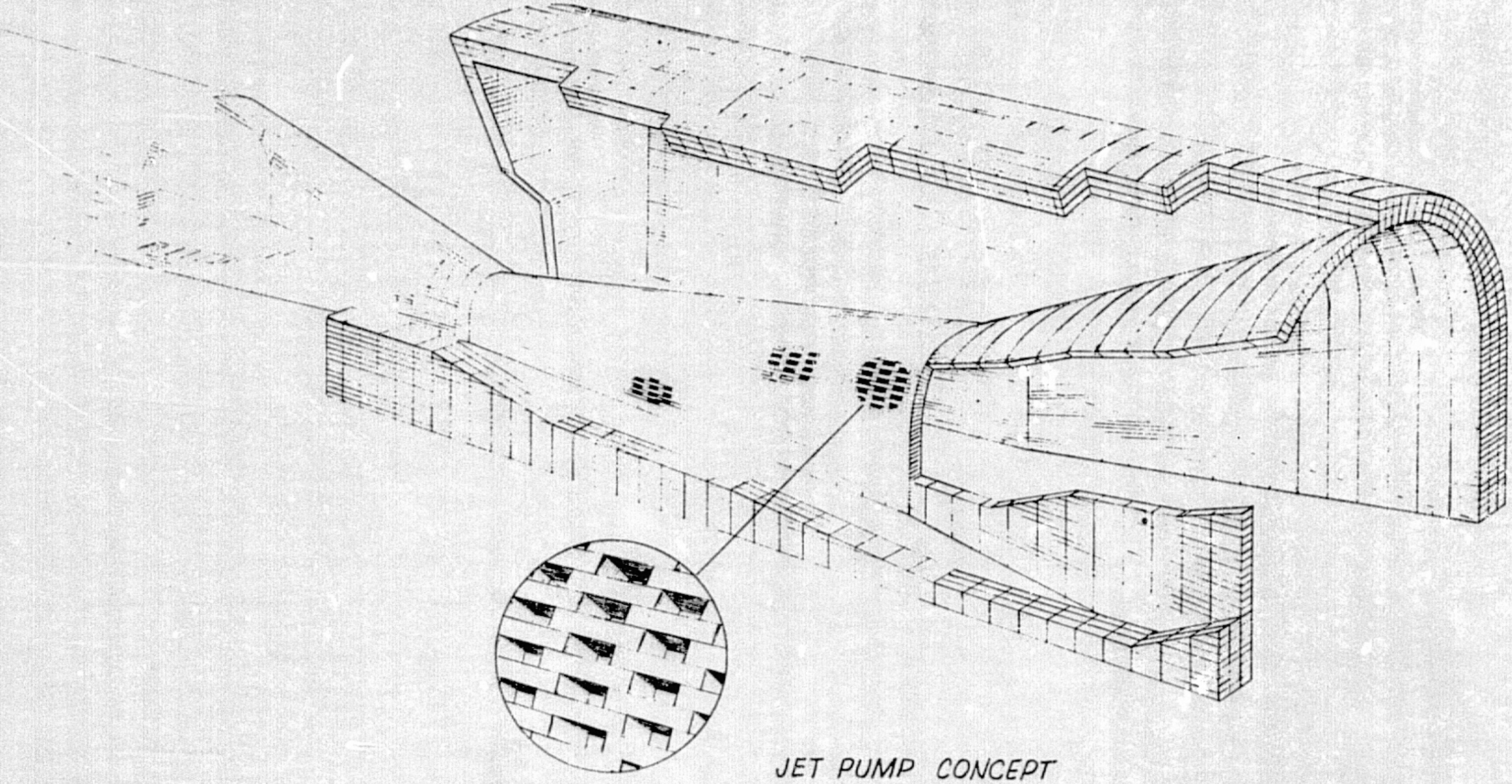


Fig. A 9.5.8—Conceptual diagram of ceramic jet pump

Table A 9.5.2 Basic Cost Rates

	Magnesia Brick	Insulation Firebrick 2800°F Rating	Insulation Firebrick 2300°F Rating	Steel Sheet
Density, lb/ft^3	179	48	31	518
Material Cost	23.38 $\$/ft^3$	7.65 $\$/ft^3$	5.10 $\$/ft^3$	1.60 $/lb
Material Cost Multiplier	1.5	1.5	1.5	1.0
Labor Cost	300 $/ton	300 $/ton	300 $/ton	4.40 $/lb
Labor Cost Multiplier	1.25	1.25	1.25	1.0

outer shell, which acts as a conduit for the recirculated flow stream, is assumed to be a cylinder of the same cross-sectional shape as the other sections.

The outer shell and the nozzle are comprised of three 15.24 cm (6 in) layers of brick (consisting of an inner lining of dense magnesium brick, a middle layer of insulating firebrick rated at 1811°K (2800°F) and a final outer layer of insulating brick rated for 1533°K (2300°F). This arrangement assures heat losses corresponding to less than 2°K (3.6°F) temperature drop in the gas streams. A backing of steel sheet of thickness 0.127 cm (0.05 in) is subjected to nominal hoop stresses of less than 68.95 MPa (10,000 psi).

The final diffuser is composed of a single layer of magnesia brick, while the diffuser-inductor section is constructed of a single open-lattice layer of magnesia brick with a 25% void. Toward the larger, higher-pressure end, this section would be supported by the outer shell.

The length of the nozzle was taken as 7.22 m (23.7 ft), about twice its inlet diameter. The length of the inductor-diffuser section was assumed to be 11.39 m (37.4 ft) based on a mean radial divergence angle of 5 degrees. The length of the final diffuser was set at 8.22 m (27.0 ft) based on a 7 degree divergence angle. The length of the outer shell was taken as 19.54 m (64 ft), or approximately the combined length of the two diffuser sections.

Table 9.5.2 summarizes the costing rates which have been used, where a multiplier has been included to allow for the fact that the bricks will have to be made in special shapes.

Table A 9.5.3 gives a summary of the cost estimate of a single inductor assembly for Base Case 1 with a flow ratio of 2.02

The cost estimates for the other cases were not made in this detail. Assuming that the thickness of all components would not change, the area volume, weight, and cost of each component was assumed to vary directly with exit flow rate. Table 9.5.4 summarizes these cost estimates.

Table A 9.5.3 Detailed Cost Estimate for Base Case 1

	Area, ft^2	Thickness, in	Volume, ft^3	Weight, lb.	Mat. Cost, $\$ \times 10^{-3}$	Labor Cost, $\$ \times 10^{-3}$	Total Cost, $\$ \times 10^{-3}$
Outer Jacket							
Mag. Lining	3400	6	1700	304,000	59.45	57.00	116.45
2800°F Brick	3581	6	1788	85,600	20.50	16.08	36.58
2300°F Brick	3740	6	1870	58,000	14.30	10.82	25.12
Steel Wall	3908	0.05	16.3	8,360	13.39	36.80	50.19
Total			737.4	455,960	107.64	120.70	228.34
Nozzle							
Mag. Lining	651	6	326	58,200	11.39	10.91	22.30
2800°F Brick	684	6	342	16,400	3.92	3.08	7.00
2300°F Brick	716	6	358	11,000	2.74	2.07	4.81
Steel Wall	749	0.05	3.1	1,620	2.59	7.21	9.80
Total			1,030	87,220	20.64	23.18	43.82
Diffuser-Inductor							
Mag. Wall	921						
With 25% Open Area	691	6	346	61,900	12.10	11.60	23.70
Final Diffuser							
Mag. Wall	1254	6	628	121,100	21.91	21.02	42.93
			7,377	717,180	162.29	176.50	388.79

Table A 9.5.4 Cost Summary All Cases

Case	M_1, lb/s	$\frac{M_2}{M_1}$	Multiplier	Vol., ft^3	Wt., lb	Mat. Cost x 10^{-3},$	Labor Cost x 10^{-3},$	Total Cost x 10^{-3},$
1	314.0	2.87	1.42	10,490	1,020,000	231.00	251.00	482.00
2	189.6	2.87	.859	6,345	616,000	139.50	151.40	290.90
3	96.0	2.87	.435	3,215	312,100	70.70	76.90	147.50
12	311.0	2.02	.991	7,300	711,000	161.00	175.00	336.00
13	330.5	1.70	.885	6,540	635,000	143.70	156.10	300.00
15	314.5	2.87	1.42	10,490	1,020,000	231.00	251.00	482.00
16	314.5	2.87	1.42	10,490	1,020,000	231.00	251.00	482.00
1	314.0	2.02	1.00	7,377	717,180	162.30	176.50	338.80
2	196.5	2.02	.604	4,455	433,400	98.10	106.50	204.90
3	96.0	2.02	.306	2,258	219,600	49.82	54.05	119.10
15	314.5	2.02	1.00	7,377	717,180	162.30	176.50	138.80
16	314.5	2.02	1.00	7,377	717,180	162.30	176.50	138.80

A 9.5.5 Alternative Solutions

Because of the very low efficiency of the jet pump for high recirculation rates, some other solution must be sought. The available alternatives include:

- Accepting a lower final preheat temperature. This reduces the required temperature of the gapor combustion air and, hence, requires a lower recirculation rate of products.
- Accepting higher gapor combustion temperatures. Some materials can tolerate the higher temperatures, but they will have limited life and increase the cost of the stove.
- Cooling the recirculated products to a temperature where they can be compressed by mechanical means. In order to achieve the desired preheat temperature, the recycled products will have to be reheated to their original temperature, or the gapor combustion air will have to be heated to an even higher temperature. This solution will require considerably more heat transfer equipment, and there will be attendant pressure and thermodynamic losses.

Although sufficient effort has not been expended to determine the best solution, some combination of one and two would appear attractive.

Appendix A 9.6

PLANT ISLAND LAYOUT

A 9.6.1 High-temperature Pipes and Valves

A large quantity of air must be handled at high temperature. Detailed piping designs, therefore, have been developed in order to estimate costs and to assist the A/E in evaluating site requirements. Layout drawings of the three base case plant islands are shown in Figures A 9.6.1 (Base Case 1), A 9.6.2 (Base Case 2), and A 9.6.3 (Base Case 3). The plant components are drawn approximately to scale and give a good indication of their relative size.

A 9.6.2 Base Case 1 - Stove Piping

In Base Case 1 combustion air, preheated at the MHD diffuser exit, is further heated by stoves similar to those used in steel blast furnaces. These stoves are arranged in two groups of 20 on each side of the MHD combustor. Fuel gas from the carbonizer and combustion air heated by a muffle furnace extracting energy from the stove exhaust are used to heat half the stoves, while the other half is heating the combustion air. It is desirable to group the stoves in banks of 4 so that the number of valves is reduced and the banks can be sequenced to provide an acceptable drop of the heated combustion air temperature. The arrangement requires extremely complex valving and switching equipment. Figure A 9.6.4 shows schematically the arrangement of 5 banks of 4 stoves each. There is an identical arrangement on the opposite side of the combustor, making a total of 40 stoves. The cost estimates for the stove piping system is broken down into 15 components in the following subsections. The last digit of the subsection number refers to the circled numbers in Figure A 9.6.4.

Dwg. 257C529

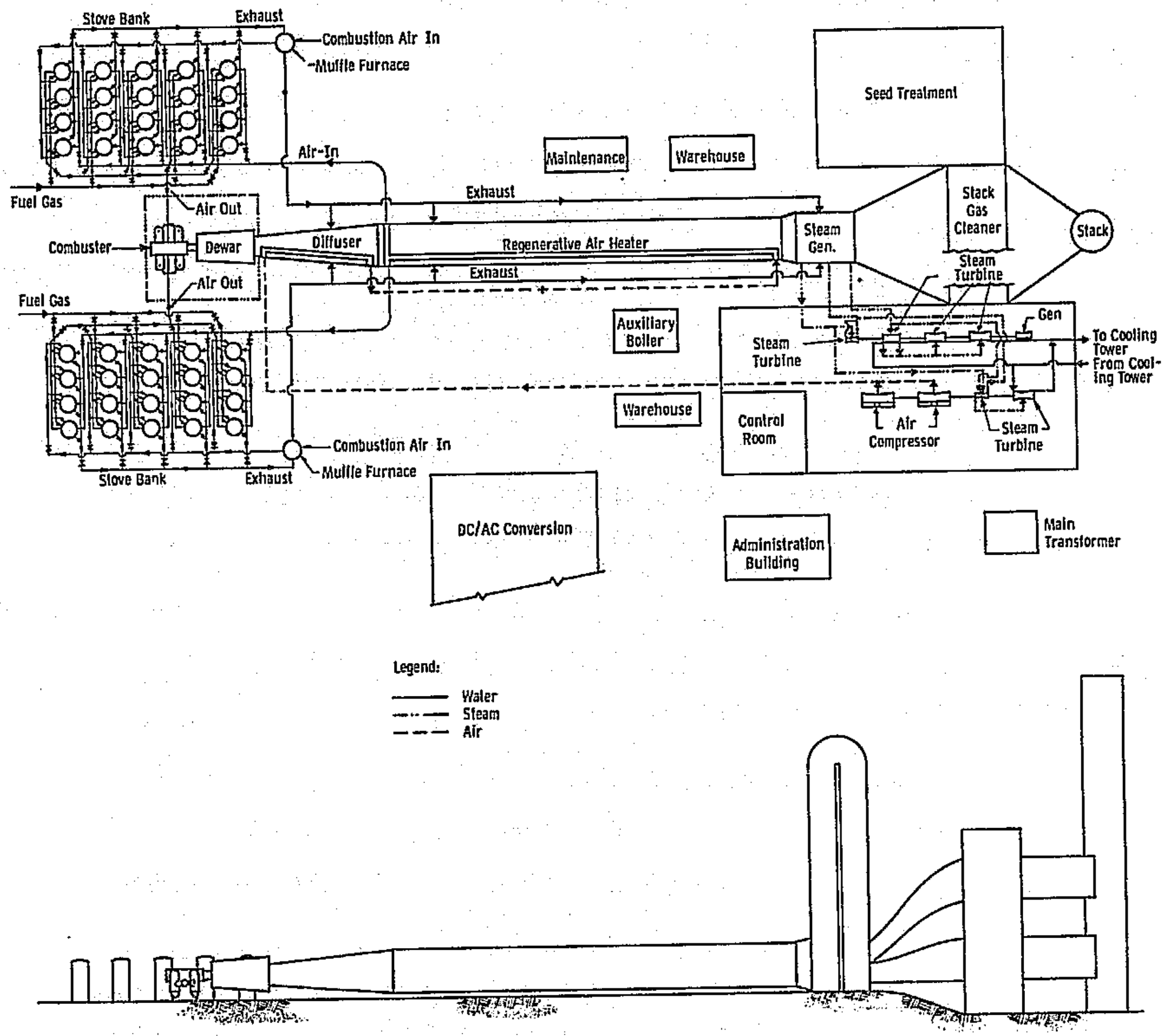


Fig. A 9.6.1—Layout of plant island for Base Case 1

Dwg. 257C530

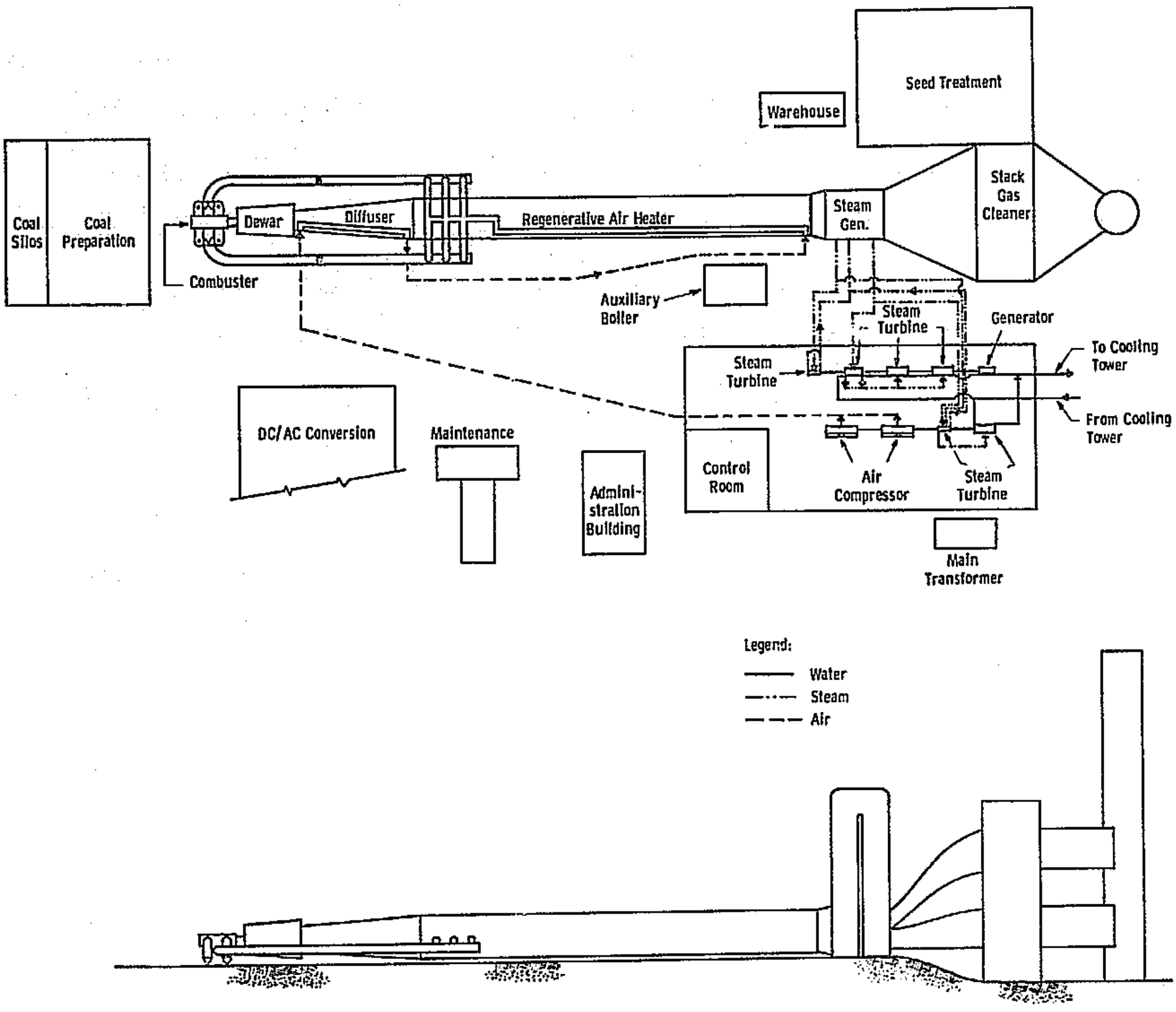


Fig. A 9.6.2—Layout of plant island for Base Case 2 (direct coal firing)

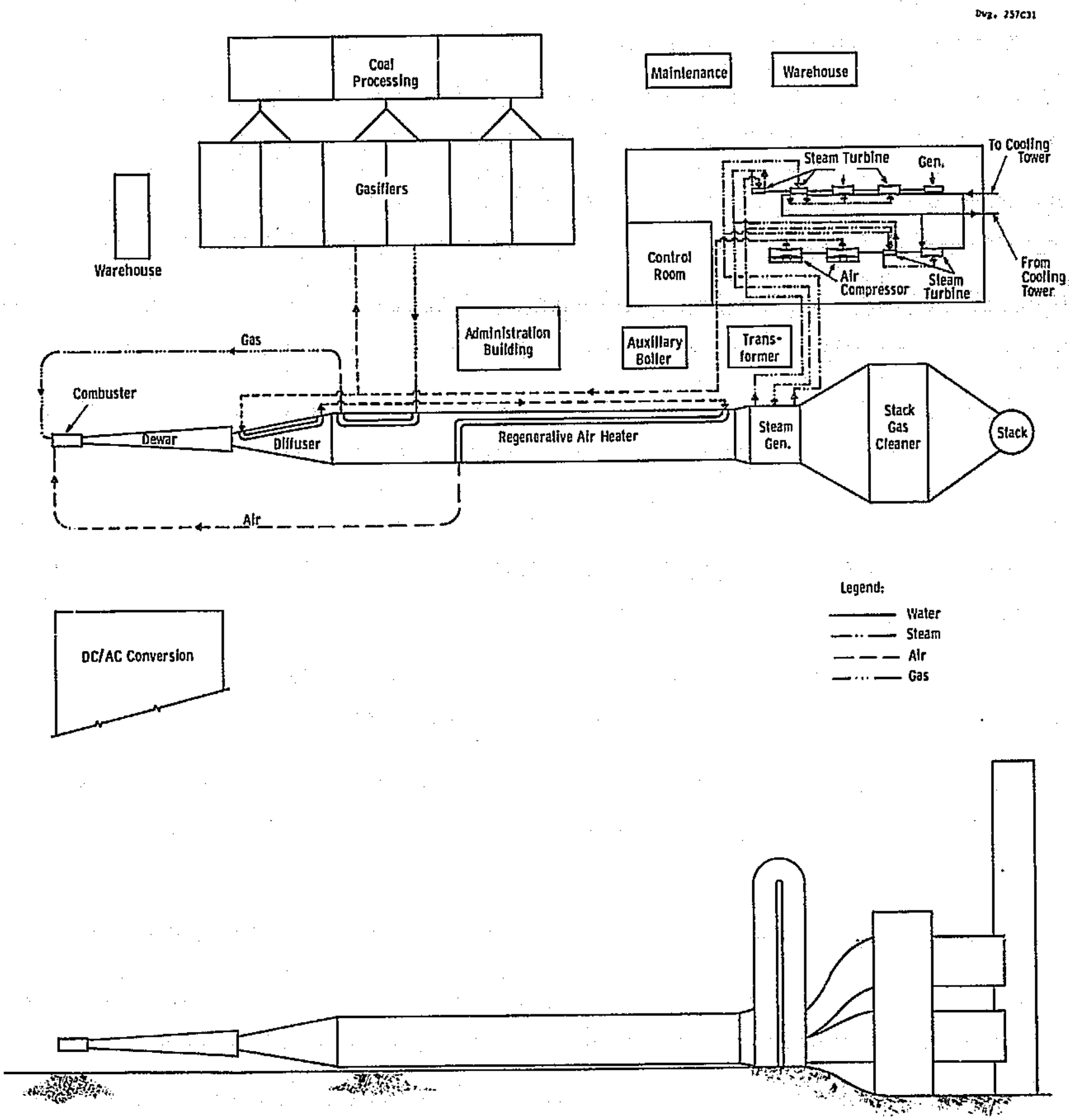
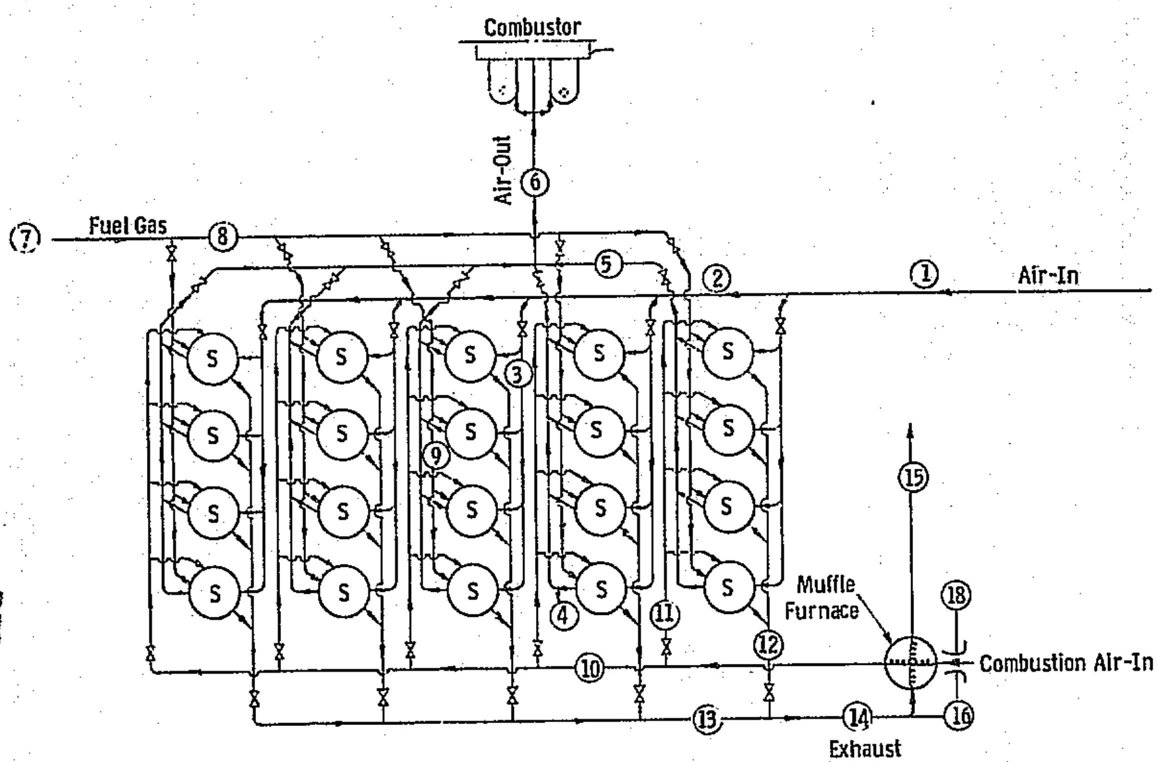


Fig. A 9.6.3--Layout of plant island for Base Case 3
(coal gasification)

Dwg. 1675B02



(1) Main Duct
(2) Main Stove Manifolds
(3) Stove Bank (Cold Blast) Manifolds
(4) Stove Bank (Hot Blast) Manifolds
(5) Stove Bank Collector Manifold
(6) Combustor Manifold
(7) Fuel Gas Pipe
(8) Fuel Gas Main Distribution Manifold
(9) Fuel Gas Stove Manifolds
(10) Combustion Air Main Distribution
(11) Combustion Air Stove Bank Manifold
(12) Stove Bank Exhaust Products Manifold
(13) Exhaust Products Main Manifold
(14) Exhaust Products to Muffle Furnace
(15) Exhaust Products to Steam Generator
(16) Exhaust Products Recycle Pipe
⋈ High Temperature Valves
(18) Recirculating Products Jet Pump

Fig. A 9.6.4—Piping diagram for separately fired air heaters

A 9.6.2.1 Preheat Air Piping

From a heat exchanger at the MHD diffuser exit, two main combustion air ducts carry heated air to the stove banks for further heating. Harbison-Walker Refractories Company was consulted for help with the duct construction design and cost estimates for the refractory portion; the U. S. Steel American Bridge Division provided data on the construction and costs for the steel pressure shell. The dimensions of these ducts are:

- An inside diameter of 3.13 m (10.28 ft) to carry the mass flow of hot air at a velocity of 73.2 m/s (240 ft/s)
- An outside diameter of 3.96 m (13 ft) to allow for 38.1 cm (15 in) of hard-faced and insulating backup brick
- A steel shell, 3.81 cm (1.5 in) thick, to contain the pressure
- High-temperature expansion seals to accomodate the length change of about 61 cm (24 in)
- Concrete, steel-reinforced, protection walls on each side of the pipe in case of rupture or spot burnout.

The length of these ducts is 67.1 m (220 ft). The combination of temperature [1588°K (2400°F)], pressure [638.2 kPa (6.3 atm)], and mass flow [1256/kg/s (2769 lb/s)] combine to make these expensive and difficult ducts to manufacture.

Table A 9.6.1 Cost Estimates for Main Air Duct

	Weight, tons/duct	Material, $/duct	Installation, $/duct
Steel Shell	202	50,500	191,900
Refractory			
Hard faced	313	56,340	76,316
Insulating backup	273	70,980	66,564
Expansion Seals	10	60,000	5,000
Concrete and Steel Supports	330	8,500	43,500
Concrete Protection Walls	462	10,000	75,000
Total per Duct	1266	256,320	485,280
Total Both Ducts	2532 tons	$512,640	$970,560

A 9.6.2.2 Main Stove Manifolds

This piping feeds air from the preheat air piping to the five stove banks. Varying the manifold diameter maintains constant header velocity in the manifold. From an initial inside diameter of 3.13 m (10.28 ft) the diameter drops to 2.80 m (9.2 ft) after the first stove bank manifold and 2.43, 1.98, and 1.4 m (7.79, 6.5, and 4.6 ft), respectively, after the other manifolds. The manifold has an overall length of 97.5 m (320 ft) of steel pipe with a 38.1 cm (15 in) thickness of refractory lining. The pipe is similar to the main preheat air pipes. Two main stove manifolds are required, and, again, estimates were based on data from Harbison-Walker and American Bridge.

Table 9.6.2 Cost Estimates: Main Stove Manifolds (2)

	Weight, tons groups	Material, \$/groups	Installation \$/groups
Steel Shell	110	27,500	95,950
Refractory			
Hard faced	202	36,360	40,500
Insulating backup	174	45,290	34,500
Expansion Seals, 4	8	160,000	15,000
Concrete & Steel Supports	50	4,000	16,000
Total	544	273,150	201,950
Total for Both Manifolds	1088 tons	\$546,300	\$403,900

A 9.6.2.3 Stove Bank Cold Blast Manifolds

The ten cold blast manifolds connect the two main stove manifolds to a total of ten banks with four stoves, half in each group. The preheated air coming from the heat exchanger at the MHD diffuser exit is fed into stove banks, where the air temperature is further raised before introducing it to the MHD combustor. Banks are switched in and out, one at a time, but sequenced so that five banks are giving up heat to the combustion air, and five banks are being heated by fuel gas at any given time.

The length of each cold blast manifold is 36.6 m (120 ft), starting with an inside diameter of 1.40 m (4.6 ft) and decreasing to a diameter of 0.7 m (2.3 ft). The same insulation thickness is required, 38.1 cm (15 in), since the air is essentially still at 1588°K (2400°F) (point 3, Figure A 9.6.4.

Table A 9.6.3 Cost Estimates for Stove Bank Cold Blast Manifold

	Weight, tons/ manifold	Material, $/ manifold	Installation, $/manifold
Steel Shell	41	10,250	35,500
Refractory			
Hard faced	22.2	3,990	4,400
Insulating backup	23.3	5,280	4,600
Expansion Seals	4.0	82,000	15,120
Concrete & Steel Supports	10.0	2,000	3,000
Total for 10 Manifolds	100.5 tons	$103,520	$62,620

A 9.6.2.4 Stove Bank Hot Blast Manifolds

After picking up heat from the stoves, the gas is collected from each stove by a hot blast manifold. Again, ten hot blast manifolds would be required for the ten stove banks. At a temperature of 1621°K (2458°F) an increase in inside diameter is required to limit the input velocity. The inside diameter begins at 0.786 m (2.58 ft), increasing with each stove until the manifold exit is 1.57 m (5.15 ft).

Table A 9.6.4 Cost Estimates for Stove Bank Hot Blast Manifold

	Weight, tons/ manifold	Material, $/ manifold	Installation, $/manifold
Steel Shell	51	12,750	12,000
Refractory			
Hard faced	62	10,670	13,000
Insulating backup	68	13,330	17,000
Expansion Seals	9	82,000	23,200
Concrete & Steel Supports	12	15,000	10,000
Total	202	133,750	75,200
Total for 10 Manifolds	2020 tons	$1,337,500	$752,000

A 9.6.2.5 Stove Bank Hot Blast Collector Manifolds

Each stove bank hot blast manifold is collected into two large manifolds leading to the combustor manifold. The collector manifold length is 117 m (385 ft). The inside diameter increases from 1.57 m (5.15 ft) to 3.51 m (11.50 ft).

Table A 9.6.5 Cost Estimates for Stove Bank Hot Blast Collector Manifolds

	Weight, tons/ manifold	Material, $/ manifold	Installation, $/manifold
Steel Shell	112	28,120	106,400
Refractory			
Hard faced	165	75,125	63,660
Insulating backup	159	72,395	61,340
Expansion Seals	14	99,000	31,000
Concrete & Steel Support	12	15,000	10,000
Total	462	289,640	272,400
Total for 2 Manifolds	924 tons	$579,280	$544,800

A 9.6.2.6. Combustor Manifolds

The collector manifolds from each group of 20 stoves carries the heated air to the combustor through two combustor manifolds. Each of these is ducted to the four combustors in such a way as to allow hot air to flow from either set of 20 stoves, or from both sets simultaneously, in varying stove bank balances. The overall manifold length is 36.6 m (120 ft), the inside diameter 3.51 m (11.5 ft) at the collector manifold attachment, and 2.47 m (8.1 ft) at the section between pairs of combustors.

Table A 9.6.6 Cost Estimates for Combustor Manifolds

	Weight, tons/ manifold	Material, $/ manifold	Installation, $/manifold
Steel Shell	20	16,000	30,000
Refractory			
Hard faced	59	13,000	13,200
Insulating backup	51	13,400	11,800
Expansion Seals	12	90,000	24,000
Concrete & Steel Support	28	38,000	32,000
Total	170	170,400	111,000
Total for 2 Manifolds	340 tons	$340,800	$222,000

A 9.6.2.7 Fuel Gas Piping

The stoves are heated by fuel gas piped from the carbonizer manifold through two main ducts to the two fuel gas distribution manifolds. These two manifolds, in turn, direct the fuel gas to ten (five from each distribution manifold) stove bank fuel gas manifolds. The overall length of the main supply ducts is 61.0 m (200 ft), and the inside diameter is 1.40 m (4.6 ft). They carry fuel gas at 811°K (1000°F) and at slightly higher than 1 atm pressure. The total mass flow of fuel gas, 45.4 kg/s (360,300 lb/hr), requires two ducts 1.40 m (4.6 ft) inside diameter at an assumed flow velocity of 36.6 m/s (120 ft/s).

Table A 9.6.7 Cost Estimates for Main Fuel Gas Supply Ducts

	Weight, tons/duct	Material, $/duct	Installation, $/duct
3/8" Steel Shell (Grade SA 516)	24	8,750	25,000
Refractory - insulating cement id, fiber-glass od	10	3,500	3,500
Expansion Seals (2)	4	11,000	3,000
Concrete & Steel Supports	10	4,000	3,500
Total	48	27,250	35,000
Total for 2 Pipes	96 tons	$54,500	$70,000

A 9.6.2.8 Fuel Gas Main Distribution Manifold

Two distribution manifolds are required whose inside diameter decreases from 1.40 m (4.6 ft) to 0.628 m (2.06 ft) as fuel gas is diverted from the main fuel gas supply ducts to stove banks. The overall length is 73.2 m (240 ft).

Table A 9.6.8 Cost Estimates for Fuel Gas Main Distribution Manifold

	Weight, tons/ manifold	Material, $/manifold	Installation, $/manifold
3/8" Steel Shell (Grade SA 516)	22	8,000	25,000
Refractory - insulating cement id, fiber-glass wrapped, od	11	3,000	3,500
Expansion Seals (5)	5	25,000	10,000
Concrete & Steel Supports	10	4,000	2,000
Total	48	40,000	40,500
Total for 2 Manifolds	96 tons	$80,000	$81,000

A 9.6.1.9 Fuel Gas Stove Manifolds

Fuel gas is carried from the main distribution manifolds to stove banks by fuel gas stove manifolds. The overall length of a manifold is 36.6 m (120 ft), the inside diameter decreasing from 0.628 m (2.06 ft), where the flow volume supplies four stoves to 0.314 m (1.03 ft), where only one stove is supplied.

Table A 9.6.9 Cost Estimates for Fuel Gas Stove Manifolds

	Weight, tons/ manifold	Material, $/ manifold	Installation, $/manifold
Steel Pipe (SA 516 3/8")	5.1	1,520	4,000
Insulation - (insulating cement and fiberglass)	2.75	750	1,500
Expansion Seals, 4	1.5	3,000	1,000
Concrete & Steel Supports	2.0	1,500	500
Pipe Flanges		500	200
Total	11.35	7,270	7,200
Total for 10 Manifolds	113.5 tons	$72,700	$72,000

A 9.6.2.10 Combustion Air, Main Distribution Manifold

This manifold carries heated air [1580°K (2384°F)] from a muffle furnace to the stove bank manifolds. The pressure is essentially 101.3 kPa (1 atm), the mass flow 253 kg/s (2,007,936 lb/hr). Two manifolds are required, their overall length 97.5 m (320 ft) with an inside diameter of 3.20 m (10.5 ft) from the muffle furnace to the first stove banks, then diminishing to 1.47 m (4.83 ft) at the fifth stove bank. Since the pressure is very low and only heated air is being carried, the construction of this manifold was assumed to be cast, reinforced concrete with a refractory lining.

Table A 9.6.10 Cost Estimates for Combustion Air Main Distribution Manifolds

	Weight, tons/ manifold	Material, $/ manifold	Installation, $/manifold
Reinforced Concrete Shell	300	30,000	50,000
Refractory			
Hard faced	106	19,170	20,500
Insulating backup	59	15,360	15,500
Site Preparation & Footers			14,000
Expansion Seal	20	70,000	15,000
Total	485	134,530	115,000
Total for 2 Manifolds	970 tons	$269,060	$230,000

A 9.6.2.11 Combustion Air Stove Bank Manifolds

Combustion air is carried to the stoves from the main distribution manifold through ten stove bank manifolds. Again, since the temperature and pressure are the same as the main distribution manifold, the construction is reinforced concrete, refractory lined. The overall length of each manifold is 36.6 m (120 ft), starting at 1.47 m (4.83 ft) id and diminishing to 0.762 m (2.5 ft) at the fourth stove. Ten manifolds are required.

Table 9.6.11 Cost Estimates for Combustion Air Stove Bank Modules

	Weight, tons/ manifold	Material, $/ manifold	Installation, $/manifold
Reinforced Concrete Shell	150	15,000	26,000
Refractory			
Hard faced	26	4,790	4,500
Insulating backup	17	3,840	3,000
Expansion Seals, 4	4	8,000	
Site Preparation & Footers			5,000
Total	197	31,630	38,500
Total for 10 Manifolds	1970 tons	$316,300	$385,000

A 9.6.2.12 <u>Exhaust Products Stove Bank Manifold</u>

The products of combustion of the fuel gas used to fire the stoves are collected in these manifolds and carried to the main collector manifold. The overall length of these manifolds is 36.6 m (120 ft) the inside diameters start at 0.808m (2.65 ft) and increase to 1.55 m (5.07 ft) at the exhaust valve.

Table A 9.6.12 Cost Estimates for the Exhaust Products Stove Bank Manifold

	Weight, tons/ manifold	Material, $/ manifold	Installation, $/manifold
Steel Shell	21	4,200	6,500
Refractory	140	26,000	28,000
Expansion Seals, 4	8	68,000	22,000
Concrete & Steel Supports	9	14,000	6,000
Total	178	112,200	62,500
Total for 10 manifolds	1780 tons	$1,122,000	$625,000

A 9.6.2.13 <u>Exhaust Products Main Collector Manifold</u>

Exhaust products from the stove bank manifolds are collected into this manifold for transport to the muffle furnace. Two manifolds are required each 97.5 m (320 ft) long. The manifold has an inside diameter starting at 1.58 m (5.18 ft) and increasing to 3.54 m (11.6 ft) at the duct to the muffle furnace.

Table A 9.6.13 Cost Estimates for the Exhaust Products Main Collector Manifold

	Weight, tons/ manifold	Material, $/ manifold	Installation, $/manifold
Steel Shell	50	14,000	45,000
Refractory			
Hard faced	220	39,000	32,000
Insulating backup	200	50,000	48,000
Expansion Seals, 4	16	160,000	40,000
Concrete & Steel Supports	___	8,000	12,000
Total	486	271,000	177,000
Total for 2 Manifolds	972 tons	$542,000	$354,000

A 9.6.2.14 Exhaust Products Main Duct to Muffle Furnace

The piping from the main exhaust products collector manifold to the muffle furnace is 3.54 m (11.6 ft) id, 30.5 m (100 ft) long. Two are required.

Table A 9.6.14 Cost Estimates for the Exhaust Products Main Duct to the Muffle Furnace

	Weight, tons/duct	Material, $/duct	Installation, $/duct
Steel Shell	20	5,000	15,000
Refractory			
Hard faced	33	6,000	6,500
Insulating backup	27	7,000	7,500
Expansion Seals	10	50,000	15,000
Concrete & Steel Supports	15	2,000	3,000
Total	105	70,000	47,000
Total for 2 Ducts	210 tons	$140,000	$94,000

A 9.6.2.15 Exhaust Products to Steam Generator Ducts

The exhaust products, after passing through the muffle furnace, are ducted to the steam generator downstream from the MHD diffuser. The length of each pipe was assumed to be 244 m (800 ft) and the inside diameter 2.32 m (7.6 ft). Two are needed.

Table A 9.6.15 Cost Estimates for the Exhaust Products to Steam Generator Ducts

	Weight, tons/duct	Material, \$/duct	Installation, \$/duct
Steel Shell	100	25,000	75,000
Insulation (fiberglass)		14,000	14,000
Expansion Seals, 2	8	22,000	8,000
Concrete & Steel Supports	40	11,500	18,500
Total	148	72,500	115,500
Total for 2 Ducts	296 tons	\$145,000	\$231,000

A 9.6.2.16 Recycled Combustion Products Piping

One pipe is required to recycle a portion of the stack gas to the compressors supplying MHD combustion air and stove bank combustion air. This pipe has an inside diameter of 3.05 m (10 ft) and is 121.9 m (400 ft) long. At 425°K, (306°F) no special insulation or refractory is required. An outer fiberglass insulation is assumed to have been used to protect personnel.

Table A 9.6.16 Cost Estimates for Recycled Combustion Products Piping

	Weight, tons	Material, $	Installation, $
Steel Shell	24	8,000	16,000
Fiberglass Wrap		1,200	2,800
Concrete & Steel Supports	10	2,200	5,300
Total	34 tons	$11,400	$24,100

A 9.6.2.17 High-Temperature Valves

The switching of stoves from heating to cool-down is done by high-temperature valves similar to those used in blast furnaces. Valving is required on air, fuel gas, combustion air, heated air, and exhaust products manifolds. In addition, blowdown valves are necessary on each stove bank to equalize pressures during switching.

Table A 9.6.17 Cost Estimates of High-temperature Valves

	Weight, tons	Material, $	Installation, $
MHD Combustion Air	80	600,000	10,000
Cold blast valves, 10	240	750,500	20,000
Hot blast valves, 10	10	1,058,000	20,000
Fuel Gas Valves, 10	50	86,000	10,000
Exhaust Products Valves, 10	80	600,000	10,000
Blowdown valves, 20	160	1,200,000	20,000
Sequencing Serve & Drive	—	340,000	85,000
Total	620 tons	$4,634,500	$175,000

A 9.6.2.18 Low-Btu Gas Piping (Base Case 3 Only)

The piping for low-Btu gas requires a length of 42.67 m (140 ft) of refractory lined steel pipe 1.89 m (6.2 ft) id. One pipe is needed.

Table A 9.6.18 Cost Estimates for Hot Low-Btu Gas Piping

	Weight, tons	Material, $	Installation, $
Steel Shell	127	32,000	135,000
Refractory			
Hard faced	84	15,200	18,000
Insulating backup	67	20,000	22,000
Expansion Seal	8	30,000	10,000
Concrete & Steel Supports	25	12,000	7,000
Protection Walls	180	28,000	8,000
Total	491 tons	$137,200	$197,000

A 9.6.7.19 Heated Air Duct (Base Case 2 Only)

Air from a heat exchanger at the exit of the MHD diffuser is carried to the MHD combustors through two ducts 2.87 m (9.4 ft) id. Harbison-Walker calculated the insulation requirements to maintain the the steel shell at 367°K (200°F) or less. Six inches of hard-faced refractory, backed up by nine inches of insulating brick is required. The overall length of the duct is 91.4 m (300 ft). A mass flow of 1256 kg/s, (9,968,254 lb/hr) at a temperature of 1589°K (2400°F), a pressure of 638 kPa (6.3 atm) and a velocity of 73.15 m/s (240 ft/s) were used to calculate the duct size. At 2.87 m (9.4 ft) id the pressure drop is 3.10 kPa (0.45 psi).

Table A 9.6.19 Cost Estimates for Heated Air Duct

	Weight, tons	Material, \$/duct	Installation, \$/duct
Steel Shell[a] ASTM 516	225	56,000	249,750
Refractory[b]			
Hard faced	396	71,280	75,000
Insulating backup	346	89,940	90,000
Refractory Anchors[b]	5.2	22,750	7,000
Concrete & Steel Supports	60.0	22,000	48,000
Expansion Skids	2	5,000	15,000
Expansion Seals[c]	2	46,000	9,000
Concrete Protection Walls	630	70,000	80,000
Total	1,666	383,220	573,750
Total for 2 Ducts	3332 tons	\$766,440	\$1,147,500

[a]Data supplied by M. Karr, American Bridge, Division, U.S. Steel.

[b]Data supplied by R. Bohac, Harbison-Walker Refractory Company.

[c]Data supplied by H. Graham, Zallea Brothers Company, Wilmington, Del.

Total weight, material, and installation cost estimates for the high-temperature piping of Base Case 1 which were given in Tables A 9.6.1 through A 9.6.17 are summarized in Table 9.22 Accounts 13.12 through 13.17.

Appendix A 9.7

LISTINGS OF COMPUTER PROGRAMS DEVELOPED FOR OPEN-CYCLE MHD CALCULATIONS

Index

<u>Functions</u>(Cont.) <u>Page</u>

1. DISKMAP

```
@RUN,M/RNPT  DPW42,U9E96FCAS40,WEI,1,50

@ASG,T  READER

@DATA,IL  READER
DATA T7 RL70-5 05/28-14:10:45
        @SFOR,IS   MAIN
        C         READ 2502 DATA AND CREATE FILE FOR DUCT PROGRAM
              IMPLICIT DOUBLE PRECISION (A-H, O-Z)
              INTEGER HEAD
              DIMENSION   ZMU(50,50), ZW(50,50), ZH(50,50), ZSIGMA(50,50),
             1      ZS(50,50), ZQADD(50,50), VP(50), VT(50),
             2      SB(10,10), SC(10,10)
              DIMENSION ZMU2(50), ZQADD2(50), ZW2(50), ZH2(50), ZSIGM2(50),
             1      ZS2(50), VT2(50)
              EQUIVALENCE  (ZMU, ZMU2), (ZQADD, ZQADD2), (ZW, ZW2), (ZH, ZH2),
             1      (ZSIGMA, ZSIGM2), (ZS, ZS2), (VT, VT2)
              DIMENSION   MVT(5), MVP(5)
              DIMENSION HEAD(24)
              READ(3, 890) HEAD
              WRITE(6, 892) HEAD
                    READ(3)        NFUEL, NAGENT, PHI
                    READ(3)        ((SC(J,I),J=1,10),I=1,NFUEL),
             1                     ((SB(J,I),J=1,10),I=1,NAGENT)
              READ(5, 800) NT, NP, NMT, NMP
              READ(5, 810) (VT(I), I=1,NT)
              READ(5, 810) (VP(I), I=1,NP)
              NNT = NMT+1
              NNP = NMP+1
              READ(5, 820) (MVT(I), I=2,NNT)
              READ(5, 820) (MVP(I), I=2,NNP)
              MVT(1) = 1
              MVP(1) = 1
              DO 10 I=1,NMT
           10       MVT(I+1) = MVT(I) + MVT(I+1)
              DO 20 I=1,NMP
           20       MVP(I+1) = MVP(I) + MVP(I+1)
              DO 40 IT=1,NMT
                    IT1 = MVT(IT)
                    IT2 = MVT(IT+1) - 1
              DO 40 IP=1,NMP
                    IP1 = MVP(IP)
                    IP2 = MVP(IP+1) - 1
                    READ(3)        ((ZMU(I,J),J=IP1,IP2), I=IT1,IT2)
                    READ(3)        ((ZQADD(I,J),J=IP1,IP2), I=IT1,IT2)
                    READ(3)        ((ZW (I,J),J=IP1,IP2), I=IT1,IT2)
                    READ(3)        ((ZH (I,J),J=IP1,IP2), I=IT1,IT2)
                    READ(3)        ((ZSIGMA(I,J),J=IP1,IP2), I=IT1,IT2)
                    READ(3)        ((ZS (I,J),J=IP1,IP2), I=IT1,IT2)
           40 CONTINUE
              WRITE(4)  HEAD
              WRITE(4)  NT, NP, NFUEL, NAGENT, PHI
              WRITE(4)  (VT(I), I=1,NT)
              WRITE(4)  (VP(I), I=1,NP)
              WRITE(4) ((SC(I,J),I=1,10),J=1,NFUEL),
             1            ((SB(I,J),I=1,10),J=1,NAGENT)
              WRITE(4)  ((ZMU (I,J), I=1,NT), J=1,NP)
              WRITE(4)  ((ZQADD(I,J), I=1,NT), J=1,NP)
              WRITE(4)  ((ZW  (I,J), I=1,NT), J=1,NP)
              WRITE(4)  ((ZH  (I,J), I=1,NT), J=1,NP)
              WRITE(4)  ((ZSIGMA(I,J), I=1,NT), J=1,NP)
              WRITE(4)  ((ZS  (I,J), I=1,NT), J=1,NP)
              WRITE(6,900) NT, NP
              WRITE(6,910) (VT(I), I=1,NT)
              WRITE(6,920) (VP(I), I=1,NP)
              DO 50 I=1,NT,5
              WRITE(6,930) I, VT(I)
              WRITE(6,940) (J, ZMU(I,J), J=1,NP,5)
              WRITE(6,940) (J, ZQADD(I,J), J=1,NP,5)
              WRITE(6,940) (J, ZW (I,J), J=1,NP,5)
              WRITE(6,940) (J, ZH (I,J), J=1,NP,5)
              WRITE(6,940) (J, ZSIGMA(I,J), J=1,NP,5)
              WRITE(6,940) (J, ZS (I,J), J=1,NP,5)
           50 CONTINUE
```

```
C             SINGLE P, MULTIPLE T
         READ(2, 890) HEAD
         READ(2) NFUEL, NAGENT, PHI2
         READ(2) ((SC(J,I),J=1,10),I=1,NFUEL),
        1             ((SB(J,I),J=1,10),I=1,NAGENT)
         READ(5, 800), NT2, NMT2
         READ(5, 810) (VT2(I), I=1,NT2)
         READ(5, 810) P2
         NNT = NMT2 + 1
         READ(5, 820)  (MVT(I), I=2,NNT)
         MVT(1) = 1
         DO 60 I=1,NMT
      60       MVT(I+1) = MVT(I) + MVT(I+1)
         DO 70 IT=1,NMT
               IT1 = MVT(IT)
               IT2 = MVT(IT) - 1
               READ(2) (ZMU2(I),I=IT1,IT2)
               READ(2) (ZWADD2(I), I=IT1,IT2)
               READ(2) (ZW2(I), I=IT1,IT2)
               READ(2) (ZH2(I), I=IT1,IT2)
               READ(2) (ZSIGM2(I), I=IT1,IT2)
               READ(2) (ZS2(I), I=IT1,IT2)
      70 CONTINUE
         WRITE(4) NT2, PHI2, P2
         WRITE(4) (VT2(I), I=1,NT2)
         WRITE(4)  (ZW2(I), I=1,NT2)
         WRITE(4)  (ZH2(I), I=1,NT2)
         WRITE(4)  (ZS2(I), I=1,NT2)
         END FILE 4
     890 FORMAT(12A6)
     892 FORMAT(1H1 12A6/ 1H , 12A6)
     800 FORMAT(4I5)
     810 FORMAT(8F10.0)
     820 FORMAT(40I2)
     900 FORMAT(1H0, 2I5)
     910 FORMAT(1H0,  10F10.3)
     920 FORMAT(1H0,  10F10.3)
     930 FORMAT(1H0,  I5, F10.3)
     940 FORMAT(1H , 5(I5,   E15.5))
         CALL EXIT
         END
END DATA.

@FIN
```

2. INPUTAFLOWS

```
@DATA,IL  READER
DATA T7 RL70-5 04/28-15:24:30
          @RUN, /RNPT JAD01,08X10PDUCS01,INPUTAFLOWS,2,100/200
          @HDG CALC. FLOWS FOR DUCT AND INPUT FOR 2502
          @FOR,IS  MAIN
      C      HF ARE HEATS OF FORMATION OF SEED COMPOUNDS
             DIMENSION AMW(4),SMW(4),HF(4)
             DIMENSION HA(24)
      C      AMW ARE THE MW OF K2,CS2 AND SMW ARE MOL. WT. OF SEED COMPOUNDS
             REAL KI1,KI3
             REAL MAMC, MAWMC,MWCMC,MWSMC, MSMC, MOXMC,MGMC,MGMCP
             DATA AMW/78.2,78.2,265.82,265.82/SMW/138.2,174.26,325.83,361.89/
            1     HF/-274960.,-338620.,-267400.,-339380./
          10 FORMAT ( )
           8 FORMAT(   12A6/12A6)
           9 FORMAT(1H1)
             WRITE(6,9)
          20 READ(5, 8,END=9999)(HA(I),I=1,24)
      C      READ IN FUEL COMPOS. AS WT. PERCENT OF CARBON C , HYDROGEN H,
      C      OXYGEN FO2, NITROGEN FN2, SULFUR S,MOISTURE,ASH
      C      HHV AND SENSIBLE HEAT OF FUEL ON AS RECEIVED BASIS
             READ(5,10) C, H, FO2, FN2, S,  ,  FH2O, ASH, HHV, FSH
      C      READ PERCENT MOIST AFTER DRYING - AS FIRED
             READ(5,10)DFH2O
      C      READ IN SEED TYPE 1= K2CO3, 2 =K2SO4, 3= CS2CO3 4= CS2SO4 AND WT.,
      C      MOLES OF WATER TO SEED MOLES    IN SEED SOLUTION AND WT. PERCENT
      C      OF SEED IN PRODUCTS
             READ(5,10) LASC, RWS, R
             R=R/100.
      C      READ IN WT. PER CENT OF O2 IN AIR + O2 AND MOISTURE IN AIR SUPPLY
             READ(5,10) AO2, RWA
      C      CALCULATE COMP. SUBSCRIPTS OF CHEM. FORM. DRY ASH FREE FUEL
             CI = C / 12.011
             HI = H/1.008 / CI
             FOI  = FO2 / 16. / CI
             FNI = FN2 / 14.008 / CI
             SI = S/32.066 / CI
             CI = 1.
      C      FMW IS MW OF DAF COAL
             FMW = 12.01 + 1.008*HI + 16.* FOI + 14.008* FNI + 32.066* SI
             HHVDAF =HHV*100./(100.-FH2O-ASH)
      C      CALCULATE HEATING VALUE IN KCAL/KG. MOLE
             HVPM = HHVDAF*2325.43/4184.*FMW
             HOF = HVPM - CI* 94040. - HI *0.5 * 68317. - SI * 69300.
      C      CALCULATE H2O ENTERING AS MOISTURE IN THE COAL AS MOLE RATIO TO
      C      DRY ASH FREE FUEL LAMBDA
             RLA =DFH2O /(100. -DFH2O - ASH) * FMW /18.016
      C      CALCULATE MOLE FRACTION OF SEED COMPOUND IN FUEL MIX. A3
             C = 1./(1. + RLA)
             ALF =RWS * SMW(LASC) / 18.016
             D =(1. + ALF) * C
             A3 = R*C* (FMW + RLA * 18.016 )/(AMW(LASC) +R*(D* FMW +D*RLA*
            1       18.016 - SMW(LASC) - ALF *18.016 ))
      C      CALCULATE MOLE FRACTION OF COMBUSTIBLE A1 AND WATER A2 IN FUEL
             A1 = C - D* A3
             A2 = A1* RLA + ALF*A3
             NFUEL = 3
             NOXI = 7
             PHI = 1.05
         210 FORMAT(5X,'NFUEL=',I10,5X,'NOXI=',I10,5X,'PHI=',F10.2,/)
             WRITE(6,210) NFUEL,NOXI,PHI
          11 FORMAT(///12A6/12A6)
             WRITE(6,11)(HA(I),I=1,24)
          13 FORMAT(28X,16HFUEL COMPOSITION/2X,11HCOMB. MOLES,4X,6HCARBON,5X,
            1       8HHYDROGEN,7X,6HOXYGEN,7X,8HNITROGEN,3X,6HSULFUR )
             WRITE(6,13)
      C      PRINT OUT THREE LINES OF SECOND INPUT SHEET FOR 2502
             HFW = -68317.
             WRITE(6,10) A1,CI,HI,FOI,FNI,SI,
          14 FORMAT(3X,25HCOMB. HEAT OF FORMATION= ,F9.1,3X,14HCOMB. DAF HHV=,
            1 F9.1/3X,19HMOIST. MOLE FRACT.=,F9.6,3X,23HSEED COMPOUND MOLE FRAC
            2      ,3HT.=,F9.6/3X,17HCOMB. MOL WT(WC)=,F9.4)
             WRITE(6,14) HOF,HHVDAF,A2,A3,FMW
      C      CALC. MOLES OF MOIST. BROUGHT IN WITH EACH MOLE OF DRY AIR EPS
             EPS = RWA /18.016 * 28.9703/ 100.
             A = 1. /(1.+EPS)
```

```
      B = (1. + ALF) * A
C     CALCULATE MOLS OF O2 PER MOL OF DRY AIR ROA
      AO2=AO2/100.
      ROA=(AO2-.2319)/(1.-AO2)*28.9703/32.
      E = (1. + EPS) /(1. + ROA + EPS )
C     CALC. MOLES OF SEED COMPOUND FOR EACH MOL OXIDANT B7
      B/ = R*E*A*(28.9703 + EPS*18.016 +ROA*32.)/(AMW(LASC) + R*(E*B*
     1    (28.9703    +EPS*18.016 + ROA * 32.) -SMW(LASC) -ALF*18.016))
C     CALC. DRY AIR MOLES PER OXIDANT MOLE - DAM
      DAM = E * (A - B* B7)
C     CALC. INPUTS FOR OXIDANT IN 2502
      B1 = 0.2079* DAM
      B2 = 0.7603 * DAM
      B3 = 0.0003 * DAM
      B4 = 0.0095 * DAM
      B8 =ROA * DAM
C     THE MOIST. ASSOC. WITH AIR IS SEPARATED FROM THAT WITH SEED SINCE
C     THEY ENTER IN DIFFERENT PHASES
      B5 = EPS * DAM
      B6 = ALF * B7
C     PRINT OUT NOS. NEEDED FOR 2502
      HCO2 = -94040.
      HB5 = -57760.
   15 FORMAT(23X,19HOXIDANT COMPOSITION /12H O2 MOL FRAC, 2X,11HN2 MOL F
     1RAC,2X,12HCO2 MOL FRAC,  2X,11HA MOL FRAC, 2X,14HSEED MOL FRAC   )
      WRITE(6,15)
      WRITE(6,10) B1,B2,B3,B4,B7
   16 FORMAT(3X,22HMOL FRACT OF VAP H2O= ,F9.6,5X,22HMOL FRACT OF LIQ H2
     10= ,F9.6/3X,23HMOL FRAC OF O2 ENRICH.= ,F9.6)
      WRITE(6,16) B5,B6,B8
C     BEGIN CALC. OF MASS FLOW RATE RATIOS
C     XS IS THE STOICHIOMETRIC MOLES OF OXIDANT PER MOLE OF FUEL
C     FIRST XS GIVES O2 REQUIRED BY COMB. + CARBONATE SEED IN FUEL+ OXID
      SM = 0.
      IF (MOD(LASC,2).EQ.0.) SM = 0.5
      XS=(A1*(CI + 0.25*HI - 0.5*FOI + SI) - SM*A3)/(B1 + SM*B7)
C     CALC. MOLECULAR WT OF OXIDANT   OMW
            OMW = DAM   *28.9703 + (B5 +B6)* 18.016 + B7*SMW(LASC)+B8*32
   17 FORMAT(3X,22HMOL FRAC OF  AIR (NA)=,F9.6,3X,19HMOL FRAC OF O2 (NOA
     1      ,2H)=,F9.4)
      WRITE(6,17) DAM,B8
C     RMRP IS THE  MASS RATIO OF RECIRC. PRODS. TO DUCT FLOW
      READ(5,10) NPHI, RMRP
      DO 100 I =1,NPHI
C     CALCULATE MASS FLOW RATIOS FOR INPUT EQUIV. RATIOS
      READ(5,10) PHI
C     RMAP IS RATIO OF SUPPLEMENTARY AIR TO MG
      RMAP =0.
C     CALCULATE MOLES OF SEED SUPPLIED FOR EACH MOLE OF DAF COMBUSTIBLE
      SMCM =(A3 + XS*PHI* B7) /A1
      MWSMC = SMCM *RWS *18.016 /FMW
      MSMC = SMCM *SMW(LASC)/FMW
      MAMC= DAM*28.9703   /FMW * XS* PHI /A1
      MWCMC = RLA*18.016/FMW
C      MOXMC MAY NOT BE CORRECT RMRP*RMAP>0.   2/20/75   DQH
      MOXMC=ROA*32./28.9703 * MAMC
      MAWMC =(1.+ RWA/100.)* MAMC
      MGMCP=(1.+ MAWMC + MWCMC + MWSMC + MSMC + MOXMC)/(1.-RMRP)
      IF (PHI .LT. 1.) RMAP=(1.05-PHI)/PHI* MAWMC/MGMCP/(1.-RMRP)
      DO 40 K=1,15
      RMAW=MAWMC/MGMCP
      MAWMC =(1.+ RWA/100.)* MAMC    - RMRP*RMAP*MGMCP
      MGMC =(1.+ MAWMC + MWCMC + MWSMC + MSMC + MOXMC)/(1.-RMRP)
      IF (ABS((MGMCP-MGMC)/MGMC).LE..0001) GO TO 42
      WRITE(6,10)MGMC,MAWMC
   40 MGMCP=MGMC
   41 FORMAT(1H0,'MGMC ITERATION HAS NOT CLOSED')
      WRITE(6,41)
   42 RMC=1./MGMC
      RMAW=RMC*MAWMC
      RMS =RMC*MSMC
      RMWC=RMC*MWCMC
      RMWS=RMC*MWSMC
      RMCW=RMC+RMWC
      RMOX=RMC*MOXMC
   22 FORMAT(25X,21HMASS FLOW RATE RATIOS/3X,4HPHI=,F7.3,5X,4HXSS=,F9.5)
```

```
      WRITE(6,22) PHI,AS
      WRITE(6,23)
   23 FORMAT(5X,4HRMOX,9X,4HRMRP,9X,4HRMAP)
      WRITE(6,10)RMOX,RMRP,RMAP
   24 FORMAT(5X,4HRMAW,9X,3HRMC,10X,3HRMS,9X,4HRMWC,10X,4HRMWS,9X,4HRMCW
     1      )
      WRITE(6,24)
  100 WRITE(6,10) RMAW, RMC, RMS, RMWC, RMWS, RMCW
      PUNCH 11,(HA(I),I=1,24)
  200 FORMAT(2I2,F10.2)
      PUNCH 200,NFUEL,NOXI,PHI
  220 FORMAT(5F14.5)
      PUNCH 220,A1,CI,HI,FOI,FNI
      AI1 = 0.0
      CSI1= 0.0
      KI1 = 0
      PUNCH 220,SI,AI1,CSI1,KI1,HOF
      CI2 = 0.0
      HI2 = 2.0
      FOI2 = 1.0
      FNI2 = 0.0
      PUNCH 220,A2,CI2,HI2,FOI2,FNI2
      SI2 = 0.0
      HOV2 = -68317.0
      PUNCH 220,SI2,AI1,CSI1,KI1,HOV2
      IF(LASC .EQ.1 .OR. LASC .EQ. 3)  CI3 = 1.0
      IF(LASC .NE.1 .AND. LASC .NE. 3)  CI3 = 0.0
      HI3 = 0.0
      IF(CI3 .EQ. 1.0  )   FOI3 = 3
      IF(CI3 .NE. 1.0  )   FOI3 = 4
      PUNCH 220,A3,CI3,HI3,FOI3,FNI2
      IF(CI3 .EQ.0.0)      SI3 = 1.0
      IF(CI3 .NE.0.0)      SI3 = 0.0
      IF(LASC .EQ. 3 .OR. LASC .EQ. 4)  CSI3 = 2.0
      IF(LASC .NE. 3 .AND. LASC .NE. 4)  CSI3 = 0.0
      IF(CSI3 .EQ. 0.0)  KI3 = 2
      IF(CSI3 .NE. 0.0)  KI3 = 0
      PUNCH 220,SI3,AI1,CSI3,KI3,HF(LASC)
      FOI4= 2.0
      FOI5 = 0.0
      DUM1 = 0.0
      PUNCH 220,B1,CI2,HI3,FOI4,FNI2
      PUNCH 220,SI2,AI1,CSI1,KI1,DUM1
      FNI5= 2.0
      DUM2= 0.0
      PUNCH 220,B2,CI2,HI3,FOI5,FNI5
      PUNCH 220,SI2,AI1,CSI1,KI1,DUM1
      CI6 = 1.0
      FOI6= 2.0
      DUM3 = -94040.0
      PUNCH 220,B3,CI6,HI3,FOI6,FNI2
      PUNCH 220,SI2,AI1,CSI1,KI1,DUM3
      AI7 = 1.0
      PUNCH 220,B4,CI2,HI3,FOI5,FNI2
      PUNCH 220,SI2,AI7,CSI1,KI1,DUM1
      DUM5 = -57760.0
      PUNCH 220,B5,CI2,HI2,FOI2,FNI2
      PUNCH 220,SI2,AI1,CSI1,KI1,DUM5
      PUNCH 220,B6,CI2,HI2,FOI2,FNI2
      PUNCH 220,SI2,AI1,CSI1,KI1,HOV2
      PUNCH 220,B7,CI3,HI3,FOI3,FNI2
      PUNCH 220,SI3,AI1,CSI3,KI3,HF(LASC)
      GO TO 20
 9999 CONTINUE
      STOP
      END
WMAP,SIX
  LIB    LIBRARY*FORTRAN
WHDG,P  TEST CASE
WXQT
      PROD. OF PPS COAL GAS - LOW BTU GAS
      AIR WITH .639% MOIST. IS OXIDANT
29.30,5.03,46.61,19.06,0.,0.,0.,5960.77,0.
0.
1,0.,0
23,19,0,639
```

3. CHRDUC2

```
I.MHDDUCT.CHRDUC2
      IMPLICIT DOUBLE PRECISION (A-H, O-Z)
      INTEGER HEAD
      EXTERNAL FUNCT2,  PWRE,FUNCT3
      DOUBLE PRECISION NA, NC, KO, MU, N, MUB, KP, K, NG,
     1       LMBDA1, LMBDA2
      COMMON        NT, NP, HEAD(24),
     1     VI(50), VP(50), ZMU(50,50), ZW(50,50), ZH(50,50),
     2     ZSIGMA(50,50), ZS(50,50), ZQADD(50,50), SC(10,10), SB(10,10)
      DIMENSION VBW(50),VBS(50)
 2109 FORMAT('0','JET PUMP EFFICIENCY=',F8.6)
 2111 FORMAT(I2,I8,F10.0)
 2113 FORMAT('0','AIR HEATR.RECIRC.RATIO=',F6.4,
     1      ' RATIO OF COAL TO CHAR HEAT VALUE=',F7.4)
      DIMENSION TI(30),HH(30),RHORHO(30),SS(30),TTRY(20),RTRY(20),X(36),
     1     RO(36),P(36),T(36),SIGMA(36),H(36),RHO(36),MU(36),U(36),
     2     S(36),N(36),B(36),MUB(36),VK(36)
      DIMENSION   DWG(7), WS(7), PS(7), PSE(7), PO(7), ETA(7)
      DIMENSION PRWT(31),PRAT(31),PROXT(31)
      COMMON /COMT2/ HAT(31), HWT(31),VTHWT(31),RWA,RWAP1,H2,HOXT(31),
     1      VBH(50),RMOX,RMAW,RMRP,VBT(50),NBT
      COMMON /REPR/HSTACK,TSTACK,PP(30),HRE
      DIMENSION LTAB1(10), LTAB2(24)
      DIMENSION    PSETAG(7), PSEPAU(7), CETA(7), PSEMHD(7), VETAS(7)
      DATA  PRWT  /     .220900,  .541200,  1.130500,  2.12000,  3.68300,
     1     6.03800,  9.46900,  1.432701,  2.10501,   3.01901,  4.24101,
     2     5.85301,  7.95301,  1.06670 2,  1.41350 2,  1.85360 2,  2.40802,
     3     3.10002,  3.96002,  5.02202,  6.32402,  7.91502,  9.84802,
     4     1.21840 3,  1.49990 3,  1.83560 3,  2.23703,  2.71303,  3.27603,
     5     3.94003,  4.72003 /
      DATA  PRAT  /     .485800,  1.059000,  2.00500,  3.44600,  5.52600,
     1     8.41100,  1.229801,  1.741301,  2.40101,  3.169400,  4.28800,
     2     5.58601,  7.17301,  9.09501,  1.140302,  1.415102,  1.740002,
     3     2.12102,  2.56602,  3.08102,  3.67602,  4.35702,  5.13502,
     4     6.01902,  7.02002,  8.14802,  9.41402,  1.083403,  1.241703,
     5     1.418003,  1.613203  /
      DATA  PROXT /  1.830500,    4.00500,  7.62900,  1.324401,  2.15301,
     1     3.33001,  4.95501,   7.14601,  1.003602,  1.378002,  1.855402,
     2     2.45502,  3.20002,   4.11402,  5.22403,  6.56002,  8.15102,
     3     1.003803,  1.225203,  1.483903,  1.784403,  2.13103,  2.53003,
     4     2.98503,  3.50403,  4.09203,  4.75603,  5.50303,  6.34403,
     5     7.28203,  8.33003  /
      DATA LTAB1/3H1  ,3H2  ,3H3  ,3H4  ,3H5  ,3H6  ,3H7  ,3H8  ,
     1     3H9  ,3H10  /
     2     LTAB2 / 3H400, 3H401, 3H402, 3H403, 3H404, 3H405, 3H406,
     3             3H407, 3H408, 3H409, 3H410, 3H411, 3H412, 3H413,
     4             3H414, 3H415, 3H416, 3H417, 3H418, 3H419, 3H420,
     5             3H421, 3H422, 3H423 /,
     6     LAB5, LAB6 / 3H5  , 3H6   /,    ZERO / 0.0D0 /
      IN = 5
      IOUT = 6
      NZDIM = 50
      TEMP = 2.22222222222D2
      DO 10 I=1,31
         VTHWT(I) = TEMP
         TEMP = TEMP + 5.55555555555D1
   10 CONTINUE
```

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```
      TOLBIS = 1.0E-3
      TAMB = 288.3
      PAMB = 1.
      RWA  = .00639
      ETAC = 0.9
      WA = 28.970
      TSTACK = 0.0
      DPIN = .02
      DPCO = 0.05
      DPAP = 0.03
      DPSG = 0.06
      DPPR = 0.06
      DPINJ = 0.0
      DB = 2.0
      KO = 0.82
      BO = 6.0
      C = 0.10
      LMBDA1 = 0.05
      LMBDA2 = 0.1
      ETAD = 0.8
      F = 0.005
      ETAG = 0.984
      ETAI = 0.985
      PAUXP1 = 0.015
      DHA = 0.0
      RWAP1 = RWA + 1.0
      KP = 101325.0
      R = 8314.69/KP
      RLH = 1.0 + LMBDA2
      CALL SETUP(PHI)
      READ(IN,2111)NUMBER
      DO 9000 NUMBE=1,NUMBER,1
      ITERAT=0
      CALL PRELIM(RMAW,RMC,RMS,RMWC,RMWS,RMCW,RMAP,RMRP,RMOX,
     1        HHVDAF,XS,WC,NC,NA,ITERAT)
      READ(IN,2100)PHIIN,PTOT,TCOMB,PCOMB,UO,TSTACK,RECIRC
      IF(PHI  .EQ.  PHIIN) GO TO 102
      WRITE(IOUT,3100)
 3100 FORMAT(//,35H PHI AND PHIIN ARE NOT CONSISTENT    ,//,
     1        38H CHECK IF PROPER DISK  FILE IS INPUT     ,//
     2       ,20H RUN TERMINATED ****       )
      STOP
  102 CONTINUE
      READ(IN,2100) RMGMC,RMAMG,HEXH,TAP,TCR,RCOCHH
      WRITE(IOUT, 2200) HEAD
      WRITE(IOUT,2210) PHIIN,PTOT,TCOMB,PCOMB,UO,RMRP,RMOX
      WRITE(IOUT,2113)RECIRC,RCOCHH
      WRITE(IOUT, 2220) RMAW, RMC, RMS, RMWC, RMWS, RMCW, RMAP
      WRITE(IOUT, 2230) WC, NC, NA, HHVDAF, XS, TSTACK
      WRITE(IOUT,2260) RMGMC,RMAMG,HEXH,TAP,TCR
      READ(IN,2111)N8T,ITERAT,TPRE
      READ(IN, 2121) (V8T(I), I=1,N8T)
      READ(IN, 2121) (V8W(I), I=1,N8T)
      READ(IN, 2121) (V8H(I), I=1,N8T)
      READ(IN, 2121) (V8S(I), I=1,N8T)
 2121 FORMAT()
  104 ITERAT=ITERAT+1
```

```
      IF(ITERAT.GT.10)GO TO 460
      IF(RMAP .GT. 1.E-30)DPINJ = .03
      HSTACK=SINTPA(VBT,VBH,TSTACK,NBT)
      RMGP = RMAP + 1.0
      RMA = RMAW / RWAP1
      TT(4) = TCOMB
      PP(4) = PCOMB
      PP(1) = PAMB*(1.-DPIN)
      PP(10) = PAMB
      TT(10) = TSTACK
      HH(10) = HSTACK
      PP(3) = (1.0+DPCO) * PP(4)
      PP(2) = (1.0+DPAP) * PP(3)
      PP(9) = (1.0+DPSG) * PP(10)
      PP(8) = (1.0+DPPR) * PP(9)
      PP(7) = (1.0+DPINJ) * PP(8)
      THETA=2325.98*HHVDAF*RCOCHH
      DRP31 = (PP(2)-PP(4)) / PP(4)
C     ***             T1 CALCULATION
      T2=TAMB
C     H IS ENTHALPY OF MIXTURE PER POUND OF MOIST AIR
      HAMB =(SINTPA(VTHWT,HAT,T2,31)+RWA*SINTPA(VTHWT,HWT,T2,31))/RWAP1
C     PRWT,PRAT AND PROX ARE THE RELATIVE PRESS. FOR WATER VAPOR, AIR
C     AND OXYGEN TAKEN FROM GAS TABLES STARTING AT 400R(222.22K) WITH
C     100R(55.556K) INCREMENTS
      PRA= SINTPA(VTHWT,PRAT,TAMB,31)*PP(2)/PP(1)
      PRW= SINTPA(VTHWT,PRWT,TAMB,31)*PP(2)/PP(1)
      DHAW=(SINTPA(PRAT,HAT,PRA,31)+RWA*SINTPA(PRWT,HWT,PRW,31))/RWAP1
     1     - HAMB
      PROX =SINTPA(VTHWT,PROXT,TAMB,31)*PP(2)/PP(1)
      DHOX=(SINTPA(PROXT,HOXT,PROX,31)- SINTPA(VTHWT,HOXT,TAMB,31))
      PCMAWR=(DHAW+RMOX/RMAW*DHOX)*2325.98/ETAC+RMRP/RMAW*PWRE(PP(10))
      HH(1)=HAMB+SINTPA(VTHWT,HOXT,T2,31)*RMOX/RMAW + HSTACK/2325.98*
     1        RMRP/RMAW
      H2 = HH(1)
      TT(1) = FUNCT3(HH(1))
      CADJ=.996
      ETAJET=0.15
      PP(12)=PP(7)/.995
      PP(15)=PP(12)/.98
      PP(14)=PP(15)/.98
      PP(19)=PP(12)*CADJ
      W2=RMAMG*RMGMC*RMC*1000/RMAW
      W5=W2*RECIRC
      W3=W2+W5
      CONST=273.15*1.2929E-3
      CONV=1.0325E6
      RHO9=CONST/288.15
      RHO2=CONST/1580
      RHO5=CONST*30.4/(28.8*1700)
      RHO3=1/(W5/(RHO5*W3)+W2/(RHO2*W3))
      POWCIR=(PP(14)-1.)*CONV*W2*(1.E-7)/(ETAC*RHO2)
      POWJET=(PP(19)-PP(15))*CONV*W5*(1.E-7)/(ETAJET*RHO5)
      PCMAWR=PCMAWR+POWCIR+POWJET
      HH(2) = HH(1)+PCMAWR/2325.98
      H2 = HH(2)
      TT(2)=FUNCT3(HH(2))
```

```
C     ***          T2 CALCULATION
                QADD = ZTP(ZQADD, TT(4), PP(4), NZDIM, NT, NP, VT, VP)
     1
      T2 = 298.16
C     H IS ENTHALPY OF MIXTURE PER POUND OF MOIST AIR
      HCOMB=ZTP(ZH,TT(4),PP(4),NZDIM,NT,NP,VT,VP)
      HAP = (SINTPA(VTHWT,HAT,TSTACK,31)+RWA*SINTPA(VTHWT,HWT,TSTACK,31)
     1     ) / RWAP1
      HAO = (SINTPA(VTHWT,HAT,T2,31)+ RWA* SINTPA(VTHWT,HWT,T2,31))/
     1     RWAP1
      QADDP = 0.4536*WC*WC*(QADD/WC/WC+LMBDA1*THETA)/1055.0/RWAP1/NA/
     1     XS/PHI/WA
      HA = HAO + QADDP
      HH(3) = HAO+(1.0+RMRP*RMAP/RMAW)*QADDP-RMRP*RMAP/RMAW*(HA-HAP)+
     1     RMRP/RMAW*(1.0-RMAP)*HCOMB/2325.98
      H2 = HH(3)
       TT(3)=FUNCT3(H2)
      H2=(SINTPA(VTHWT,HAT,TCR,31)+RWA*(SINTPA(VTHWT,HGT,TCR,31)))/
     1     RWAP1+SINTPA(VBT,VBH,TCR,NBT)*RMRP/RMAW/2325.98
                QAMAWR = (H2 - HH(2)) * 2325.98
                HH(4) = ZTP(ZH, TT(4), PP(4), NZDIM, NT, NP, VT, VP)
                SS(4) = ZTP(ZS, TT(4), PP(4), NZDIM, NT, NP, VT, VP)
                RHORHO(4) = ZTP(ZW, TT(4), PP(4), NZDIM, NT, NP, VT, VP) *
     1                    PP(4)/(R*TT(4))
                RO(1) = 1.0
                X(1) = 0.0
                S(1) = SS(4)
                SS(5) = SS(4)
      U(1) = U0
                HH(5) = HH(4) - 0.5*U(1)*U(1)
                H(1) = HH(5)
                TT(5) = TZZ(S(1), ZS, H(1), ZH, NZDIM, NT, NP, VT, VP)
                T(1) = TT(5)
                P6T = PZT(S(1), ZS, T(1), NZDIM, NT, NP, VT, VP)
                PP(5) = P6T
                P(1) = P6T
                SIGMA(1) = ZTP(ZSIGMA, T(1), P(1), NZDIM, NT, NP, VT, VP)
                RHORHO(5) = ZTP(ZW, T(1), P(1), NZDIM, NT, NP, VT, VP) *
     1                    P(1)/(R*T(1))
                RHO(1) = RHORHO(5)
                MU(1) = ZTP(ZMU, T(1), P(1), NZDIM, NT, NP, VT, VP)
      PE=PTOT*.7
  170 ISAFE = 6
      PMHD = ETAI * PE
      K = K0
      DK = 0.05 - 0.0025 * PMHD * 1.0E-8
      VK(1) = K
      B(1) = B0
                MUB(1) = B(1) * MU(1)
                JSAFE = 5
                P5S = PP(7) + 0.08
                MG = PE / ((ZPZ(ZH, P5S, SS(4), ZS, NZDIM, NT, NP, VT, VP)
     1               - HH(4)) * (0.05 - K))
      IF(MG .LE. 0.0)  MG = 1300.0 * PMHD / 1.3E9
                KSAFE = JSAFE
  201           CONTINUE
                     DN = 0.0
```

```
                    IY = 1
                    IQ = 0
                    D4 = DSQRT(MG/(U(1)*RHO(1)))
                    RD4 = 1.0/D4
                    FXI = RHO(1)*U(1)*F*2.0/(B(1)**2*SIGMA(1)*D4)
                    N(1) = K*(1.0-K)/(1.0-K+FXI)
                    IV = 10
                    DP = 10.0
                    DP6T = 10.0
                    P6T = 10.0
                    JCNT = 11
C       ***        REPLACE XP CALL , II = XP(P(1), 0) + 2
                DO 290 I=1,NP
                IF (P(1) .GE. VP(I))   GO TO 290
                      II = I
                      GO TO 295
      290         CONTINUE
                II = NP
      295       CONTINUE
C                           P6T LOOP
      301       CONTINUE
                    IF (MOD(IQ,2) .EQ. 0)  GO TO 305
                        P6T = P6T + DP6T
                        P(IY) = (P6T-PP(7))/DP6T*DP + P(IY-1)
                        DP = P(IY-1) - P(IY-2)
                        DN = N(IY-1) - N(IY-2)
                        FXI = FXIM - FXI
                        GO TO 315
      305           CONTINUE
                        IY = IY+1
                        II = II-1
C       ***     REPLACE PX CALL ,  P(IY) = PX(II, 0)
                    P(IY) = VP(II)
      315           HS = ZPZ(ZH, P(IY), S(IY-1), ZS, NZDIM, NT, NP, VT,
     1                       VP)
                    DHS = HS - H(IY-1)
                    TEMP = 2.0*C*DHS + U(IY-1)**2
                    IF (TEMP .GE. 0.0D0)   GO TO 325
                        WRITE(IOUT, 2310) TEMP, C
                        GO TO 500
      325           U(IY) = DSQRT(TEMP)
                    DPOLD = DP
                    DP = P(IY) - P(IY-1)
                    N(IY) = DN/DPOLD*DP + N(IY-1)
                    FXIM = 2.0*FXI
C                       VARY FXIM, CONVERGENCE BASED ON N
                    FACTX = 8.0*KP*(1.0+C)
                    FACTN = 0.5*RLH*(1.0+C)
                    FT2 = 2.0*F
C
      351           CONTINUE
                    H(IY) =   (FACTN*(N(IY)+N(IY-1))-C)*DHS + H(IY-1)
                    T(IY) = TPZ(P(IY), H(IY), ZH, NZDIM, NT, NP, VT, VP)
                    RHO(IY) = ZTP(ZW, T(IY), P(IY), NZDIM, NT, NP, VT, VP)
     1                      *P(IY)/(R*T(IY))
                    SIGMA(IY) = ZTP(ZSIGMA, T(IY), P(IY), NZDIM, NT, NP,
     1                      VT, VP)
```

```
                    RD(IY) = DSQRT(MG/(U(IY)*RHO(IY))) * RD4
                    K = VK(IY-1) + DK*DHS*RHO(IY)/101300.
                    VK(IY) = K
                    B(IY) = B(IY-1)
                    IF (P(IY) .LE. 2.0)  B(IY) = B(IY) +
         1               DB*DHS*RHO(IY)/101300.0
                    FXIOLD = FXI
                    FXI = FT2*RHO(IY)*U(IY) / (SIGMA(IY)*RD(IY)*D4*
         1                    B(IY)**2)
                    FXIM = FXIM - FXIOLD + FXI
                    OLDNY = N(IY)
                    N(IY) = K*(1.0-K)/(1.0-K+FXI)
                    IF (DABS(OLDNY-N(IY)) .LT. 1.0E-5)     GO TO 355
                    JCNT = JCNT -1
                    IF (JCNT .LE. 1)  GO TO 355
              GO TO 351
      355           X(IY) = DP / ((K-0.5*FXIM-1.0) * (SIGMA(IY-1)+SIGMA(IY
         1              )) * (B(IY-1)**2 + B(IY)**2) * (U(IY-1)+U(IY))) *
         2               FACTX + X(IY-1)
                    S(IY) = ZTP(ZS, T(IY), P(IY), NZDIM, NT, NP, VT, VP)
                    MU(IY) = ZTP(ZMU, T(IY), P(IY), NZDIM, NT, NP, VT, VP)
                    MUB(IY) = MU(IY)*B(IY)
                    DN = N(IY) - N(IY-1)
                    JCNT = 4
                    IF (P(IY) .GT. PP(7))    GO TO 365
                         HH(7) = 0.5*U(IY)**2 + H(IY)
                         TSS = TZZ(S(IY), ZS, HH(7), ZH, NZDIM, NT, NP,
         1                          VT, VP)
                         PSS = PZT(S(IY), ZS, TSS, NZDIM, NT, NP, VT, VP)
                         P6TOLD = P6T
                         P6T = ETAD*(PSS-P(IY)) + P(IY)
                         DP6T = P6TOLD - P6T
      365           Q = IQ
                    IQ = 0
                    IF ((P6T-10.0*Q) .LE. PP(7))  IQ = 1
    C                    CONVERGENCE TEST FOR P6T
                    IV = IV-IQ
                    IF (IV .LT. 0)  GO TO 375
                    IF (DABS(DP6T) .LT. 1.0D-10)  GO TO 375
                    IF (DABS(PP(7)-P6T) .LT. 0.005)  GO TO 375
              GO TO 301
      375 CONTINUE
                    OLDMG = MG
                    MG = PE*RLH/(HH(4)-HH(7))
                    IF (DABS(OLDMG/MG-1.0) .LT. 2.0D-4)  GO TO 405
                    KSAFE = KSAFE - 1
                    IF (KSAFE .LE. 0)  GO TO 405
              GO TO 201
    C                    END OF KSAFE (FXI) LOOP
      405 CONTINUE
                HH(6) = H(IY)
                SS(6) = S(IY)
                PP(6) = P(IY)
                TT(6) = T(IY)
                RHORHO(6) =  RHO(IY)
                H5SP = ZPZ(ZH, P55, SS(4), ZS, NZDIM, NT, NP, VT, VP)
                ETAMHD = PE / (MG*(HH(4)-H5SP))
```

```
                 TT(7) = TPZ(PP(7), HH(7), ZH, NZDIM, NT, NP, VT, VP)
                 SS(7) = ZTP(ZS, TT(7), PP(7), NZDIM, NT, NP, VT, VP)
                 RHORHO(7) = ZTP(ZW, TT(7), PP(7), NZDIM, NT, NP, VT, VP) *
     1                          PP(7)/(R*TT(7))
                 TT(8) = TT(7)
                 SS(8) = SS(7)
                 RHORHO(8) = RHORHO(7)
                 HH(8) = HH(7)
                 IF (RMAP .LE. 1.0D-30)  GO TO 425
  411                  DHA =(SINTPA(VTHWT, HAT, TT(8), 31) +
     1                       SINTPA(  VTHWT,HWT,TT(8),3;)*RWA)/RWAP1-HAMB
                       HH(8) = (RMAP*DHA + HH(7)) / (RMAP+1.0)
                       TT8NEW = SINTPA(V8H, V8T, HH(8), N8T)
                       IF (DABS(TT(8)-TT8NEW) .LE. 0.1)  GO TO 415
                          TT(8) = TT8NEW
                 GO TO 411
  415                  TT(8) = TT8NEW
                       SS(8) = SINTPA(V8T, V8S, TT(8), N8T)
                       RHORHO(8) = SINTPA(V8T, V8W, TT(8), N8T) * PP(8)/
     1                          (R*TT(8))
  425            HH(9) = HH(8) - QAMAWR*RMAW/RMGP
                 TT(9) = SINTPA(V8H, V8T, HH(9), N8T)
                 SS(9) = SINTPA(V8T, V8S, TT(9), N8T)
                 RHORHO(9) = SINTPA(V8T, V8W, TT(9), N8T) * PP(9)/(R*TT(9))
           DQI = (1.0 - ETAI) * PMHD
                 QS0 = MG * RMGP * (HH(9)-HH(10))
           T2 = TCR
           HR13 = SINTPA(V8T,V8H,T2,N8T)
           HH(13) = (SINTPA(VTHWT,HAT,T2,31) + RWA * SINTPA(VTHWT,HWT,T2,31))
     1           /RWAP1 + HR13 * RMRP / RMAW * 2325.98
           T2 = TAP
           HH(15) = (SINTPA(VTHWT,HAT,T2,31) + RWA * SINTPA(VTHWT,HWT,T2,31))
     1           /RWAP1
           QS0 = QS0 + RMGMC*RMC*MG*((1.0+RMAMG)*(HEXH-HSTACK)-RMAMG*
     1           (HH(15) - HAMB))
                 QC = RMC * THETA * MG
                 QS1 = LMBDA1 * QC
                 QS2 = LMBDA2 * PE
                 PC = PCMAWR * RMAW * MG
           IF(ISAFE .LE. 0)  GO TO 448
           ETASF= 0.44*.995
           PS44 = ((1.0-ETAG)*PC - QS0 - QS1 - QS2 - DQI)*ETASF/
     1           ((1.0 / ETAG - 1.0) * ETASF-1.)-PC*ETAG
           PCAL = PS44 * (1.0 - PAUXP1) + PMHD
           IF(DABS(PCAL / PTOT - 1.0) .LE. 5.0D-3)  GO TO 448
           ISAFE = ISAFE - 1
           PE = PE * PTOT / PCAL
           GO TO 1/0
       448 CONTINUE
           TTRY(ITERAT)=TT(3)
           RTRY(ITERAT)=RMRP
           IF(ITERAT.EQ.1)GO TO 460
           TEST=ABS((TPRE-TT(3))/TT(3))
           IF(TEST.LE.0.00020)ITERAT=10
           IF(ITERAT.GE.10)GO TO 456
           IF(ITERAT-2)460,450,454
       450 RMRP=RMRP/1.2
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          GO TO 456
      454 TEST=ABS(TTRY(ITERAT)-TTRY(ITERAT-1))
          IF(TEST.LT.0.001)GO TO 456
          RMRP=RMRP+(RMRP-RMRY(ITERAT-1))*(TPRE-TTRY(ITERAT))/
         1      (TTRY(ITERAT)-TTRY(ITERAT-1))
      456 CALL PRELIM(RMAW,RMC,RMS,RMWC,RMWS,RMCW,RMAP,RMRP,RMOX,
         1      HHVDAF,XS,WC,NC,NA,ITERAT)
          GO TO 104
      460 CONTINUE
          WRITE(IOUT,2300) MG
          WRITE(IOUT,2109)ETAJET
          WRITE(IOUT,2500)ETAD,ETAG,ETAI
          WRITE(IOUT,2600)LMBDA1,LMBDA2
          WRITE(IOUT,2700)F,PAUXP1,DRP31,PCHAWR,QAMAWR,THETA,PMHD,PE
          WRITE(IOUT,3000)C,DK,DB
                ETAS = 0.39
                DO 480  KY=1,7
                      ETAS = ETAS + 0.01
                      ETASF = 0.995 * ETAS
                      RETAG = 1.0/ETAG
                      VETAS(KY) = ETAS
                      PS(KY) = (((1.0-ETAG)*PC -QS0 - QS1 - QS2 - DQI)*ETASF/
                              ((RETAG-1.0)*ETASF-1.0)
                      PSE(KY) = PS(KY) - PC*ETAG
                      DQG(KY) = PSE(KY) * RETAG - PSE(KY)
                      QS(KY) = 0.995 * (DQG(KY)+QS0+QS1+QS2+DQI)
                      PO(KY) = (1.0-PAUXP1)*PSE(KY) + PMHD
                      ETA(KY) = PO(KY)/QC
      480       CONTINUE
C                     OUTPUT
                LPP3 = 10.0*PP(9)
                LK = 100.0*K
                YY = 10.0*DRP31
                IV = YY + 0.5
                KY = 10.0*(YY-IV)
                LUU = U(1)
                LC = 100.0*C + 0.5
                LTT3 = TT(9)
                LBU = B(1)
                RMRPMG = RMRP*MG
                RMOXMG = RMOX*MG
                RMAWMG =  RMAW *MG
        RMAMG = RMA*MG
                RMGPMG=  RMGP *MG
                RMAPMG = RMAP *MG
                RMCMG  = RMC *MG
                RMCWMG = RMCW *MG
                RMSMG  = RMS* MG
                RMWCMG = RMWC *MG
                RMWSMG = RMWS *MG
                RMARP = RMAWMG+RMRPMG
                RGMAMG = MG +RMAPMG
          WRITE(IOUT,2800) PP(1), TT(1),PP(2),TT(2),PP(3),TT(3)
          WRITE(IOUT,2900) PP(4),TT(4), PP(7),TT(7),PP(8),TT(8),
         1      PP(9),TT(9), PP(10),TT(10)
                WRITE(IOUT, 2920)
          WRITE(IOUT,2424)PAMB,TAMB,HAMB,RMAWMG
```

```
      WRITE(IOUT,2423) LTAB1(1),PP(1),TT(1),HH(1),RMAWMG
      WRITE(IOUT,2423) LTAB1(2),PP(2),TT(2),HH(2),RMARP
      WRITE(IOUT,2423) LTAB1(3),PP(3),TT(3),HH(3),RMARP
      WRITE(IOUT,2425)(LTAB1(J),PP(J),TT(J),HH(J),RHORHO(J),SS(J),MG,
     1     J=4,6 )
      WRITE(IOUT,2425)LTAB1(7),PP(7),TT(7),HH(7),RHORHO(7),SS(7),MG
      WRITE(IOUT,2425)(LTAB1(J),PP(J),TT(J),HH(J),RHORHO(J),SS(J),
     1        RGMAMG,J=8,10)
              WRITE(IOUT, 2430)
              JY = IY - 1
              IF (JY .LE. 0)      GO TO 485
              WRITE(IOUT,2435)(LTAB2(J),X(J), RD(J), P(J), T(J), SIGMA(J),
     1        H(J), RHO(J), U(J), MU(J), S(J), N(J), MUB(J),
     2        VK(J), B(J), J=1,JY)
  485         J = IY
              WRITE(IOUT, 2435)  LAB3, X(J),RD(J), P(J), T(J), SIGMA(J),
     1            H(J), RHO(J), U(J), MU(J), S(J), N(J), MUB(J)
     2        , VK(J), B(J)
              V = 0.5 * D4 * (RD(IY)+1.0)
              XVD4 = (X(IY) + V) * V
              D5 = RD(IY) * D4
              A4 = D4 * D4
              A5 = D5*D5
              WRITE(IOUT, 2440)  MG, RMANG, RMAWMG, RMGPMG, RMAPMG,
     1            RMCMG, RMCWMG, RMSMG, RMWCMG, RMWSMG, XVD4, D4, D5,
     2            A4, A5, QC, ETAMHD
              DO 490 KY=1,7
                  PSETAG(KY) = PSE(KY)/ETAG
                  PSEMHD(KY) = PSE(KY) + PMHD
                  PSEPAU(KY) = PSE(KY) * PAUXP1
                  CETA(KY) = 3412.16/ETA(KY)
  490         CONTINUE
      WRITE(IOUT,2460)
      WRITE(IOUT,2480)(VETAS(J),J=1,7)
      WRITE(IOUT,2470)(QS0,J=1,7),(QS1,J=1,7)
     1            , (QS2,J=1,7), (DQI,J=1,7), (DQG(J),J=1,7),
     2            (QS(J),J=1,7), (PS(J),J=1,7),
     3            (PSETAG(J),J=1,7), (PC,J=1,7), (PSE(J),J=1,7),
     4            (PMHD,J=1,7), (PSEMHD(J), J=1,7), (PSEPAU(J),J=1,7),
     5        (PO(J),J=1,7),(CETA(J),J=1,7)
      WRITE(IOUT,2490)(ETA(J),J=1,7)
  499 CONTINUE
  500 CONTINUE
 2100 FORMAT(8F10.0)
 2110 FORMAT(I2)
 2120 FORMAT(8F10.0)
 2200 FORMAT(1H1, 12A6 / 1H , 12A6)
 2210 FORMAT(1H0, 6HPHI = , F6.2, 5X, 6HPTOT =, 1PE12.5, 5X, 7HTCOMB =,
     1        0PF8.1, 5X, 7HPCOMB =, F8.3, 5X, 4HOU = , F6.2,
     2        /,1X,7HRMRP = ,F10.3,5X,7HRMOX = ,F10.3)
 2220 FORMAT(1H0, 6HRMAW =, 1PE14.6, 5X, 6H RMC =,E14.6, 5X,  6H RMS =,
     1        E14.6 /1H , 6HRMWC =, E14.6, 5X, 6HRMWS =, E14.6, 5X,
     2        6HRMCW =, E14.6, 5X, 6HRMAP =, E14.6)
 2230 FORMAT(1H0, 4HRC = , F8.2, 5X, 4HNC =, F8.4, 5X, 4HNA = F8.4,
     1        5X, 8HHHVDAF =, 1PE14.6 /1H , 4HXS =, E14.6, 5X,
     2        10HT(STACK) =, E14.6)
 2240 FORMAT(1H0, 24H**ITERATION ON TI FAILED)
```

```
      2250 FORMAT(1H0, 24H**ITERATION ON T2 FAILED)
      2260 FORMAT(1H0, 7HRMGMC =, 1PE14.6, 5X,7HRMANG =,E14.6, 5X, 6HMEXH =,
     1           E14.6, 5X, 6H TAP= ,E14.6, 5X, 6H TCR =,E14.6)
      2300 FORMAT(1H , 6H    MG, 1PE13.6)
      2310 FORMAT(1H0/1H0, 42H** CALCULATION OF FLUID VELOCITY INVOLVES ,
     1           22HTAKING SQUARE ROOT OF , 1PE12.5, 1H. /1H , 5X,
     2           21HCURRENT VALUE OF C = , E12.5,
     3           43H. HIGHER VALUES OF C WILL ALSO BE IGNORED. )
      2400 FORMAT(1H , 2X, 2(1H-, I2), 1H-, 2I), 1H-, I3, 1H-, I2, 1H-, I4,
     1           1H-, I1)
      2410 FORMAT(1H0, 3X, 5HINPUT/1H0, 10X, 5H PHI , F5.2, 7H CYCLE , I2,
     1           4H P3 , F6.2, 4H T3 , F5.0,12H (P1-P3)/P3 , F6.2,
     2           3H U , F4.0, 3H C , F5.2, 9H LAMBDA1 , F5.2, 9H LAMDA2  ,
     3           F5.2 /1H0, 3H K , F5.2, 4H PE , 1PE8.1, 4H P6 , 0PF6.2,
     4           6H ETAD , F5.2, 3H F , F6.3, 4H T8 , F4.0, 3H B , F2.0,
     5           8H PC/MAW , F7.0, 4H P1 , F6.2, 4H T1 , F5.0, 4H P2 , F6.2
     6           /1H0, 4H T2 , F5.0, 6H ETAG , F6.3, 6H ETAI , F6.3,
     7           10H PAUX/PSE , F6.3, 8H QA/MAW , F7.0, 7H THETA , 1PE10.3
     8           / (1H0, 6H ETAS , 0PF6.2, 6(1H, F6.2)))
      2420 FORMAT(1H0,//,11X,          10HGAS STATES  ,//,2X,6HPOINT,5X,1HP,
     1           10X, 1HT, 10X, 1HH, 10X, 3HRHO, 9X, 1HS,9X,1HA,//)
      2425 FORMAT(1H , 2X, A3, 4X, F6.3, 4X, F5.0, 4X, F7.0, 4X, F7.4,
     1           4X, F7.1,4X,F8.3)
      2430 FORMAT(1H1,12X, 16HGENERATOR DESIGN  /1H0,7H  POINT,
     1           2X, 6HLENGTH, 4X, 4HD/D4, 5X, 1HP, 7X, 1HT, 4X, 5HSIGMA,
     2           7X, 1HH, 7X, 3HRHO, 6X, 1HU, 7X, 2HMU, 8X, 1HS, 7X, 1HN,
     3           6X, 4HMU*B, 5X, 1HK, 5X, 1HB)
      2435 FORMAT(1H , 3X, A3, 2X, F7.3, 2X, F6.3, 2X, F6.3, 2X, F5.0,
     1           2X, F6.3, 2X, F8.0, 2X, F7.4, 2X, F6.1, 2X, F7.4, 2X, F7.1,
     2           2X, F6.4, 2X, F7.4, F6.3, F6.2)
      2440 FORMAT(1H0, 3X, 25HMASSFLOWS, GENERATOR AREA
     1           /1H , 5X, 4H MG , F8.3, 4H MA , F8.3, 4H MAW, F8.3,
     2                4H MG*, F8.3, 4H MA*, F8.3
     3           /1H , 5X, 4H MC , F8.3, 4H MCW, F8.3, 4H MS , F8.3,
     4                4H MWC, F8.3, 4H MWS, F8.3
     5           /1H , 6X, 15H (X + DM) * DM , F9.5, 3H D4, F7.4,
     6                3H D5, F7.4, 3H A4, F7.4, 3H A5, F7.4
     7           /1H , 6X, 12H THETA * MC , 1PE12.5, 8H ETAMHD , 0PF8.5)
      2423 FORMAT(1H ,2X,A3,4X,F6.3,4X,F5.0,4X,F9.0,22X,4X,F8.3)
      2424 FORMAT(1H ,2X,3H0   ,4X,F6.3,4X,F5.0,4X,F9.0,22X,4X,F8.3)
      2460 FORMAT(1H0,13X,              19HOVERALL PERFORMANCE  )
      2470 FORMAT(1HH,4X,8HQSU              ,3X,7E14.5
     3           /1H , 4X, 8HQS1        ,3X,7E14.5
     4           /1H , 4X, 8HQS2        ,3X,7E14.5
     5           /1H , 4X, 8HDQI        ,3X,7E14.5
     6           /1H , 4X, 8HDQG        ,3X,7E14.5
     7           /1H , 4X, 8HWS         ,3X,7E14.5
     8           /1H , 4X, 8HPS         ,3X,7E14.5
     9           /1H , 4X, 8HPT         ,3X,7E14.5
     1           /1H , 4X, 8HPC         ,3X,7E14.5
     2           /1H , 4X, 8HPSE        ,3X,7E14.5
     3           /1H , 4X, 8HPMHD       ,3X,7E14.5
     4           /1H , 4X, 8HPSE+PMHD,3X,7E14.5
     5           /1H , 4X, 8HPAUX       ,3X,7E14.5
     6           /1H , 4X, 8HP          ,3X,7E14.5
     7           /1H , 4X, 8HHR         ,3X,7E14.5)
      2480 FORMAT(1H0    ,11HSTEAM PLANT   ,/ ,1X,10HEFFICIENCY ,7(2X,F12.2))
```

```
 2490 FORMAT(1H0    ,7HOVERALL ,/,1X,11HEFFICIENCY   ,2X,0PF12.4,
     1         6(2X,F12.4))
 2500 FORMAT(1H0,1X,12HEFFICIENCIES  /,5X,8HDIFFUSER ,12X,F10.4
     1         ,5X,16HROTATING GENERATOR ,2X,F10.4 , /
     2         ,5X,14HDC/AC INVERTER ,6X,F10.4   )
 2600 FORMAT(1H0,1X,19HHEAT TRANSFER RATIO , / , 6X,
     3         37HFROM COMBUSTER TO SUBPOSED PLANT   = , F8.3, / ,6X,
     4         37HMHD GENERATOR TO SUBPOSED PLANT    = , F8.3 )
 2700 FORMAT(//,6X,29HFRICTION FACTOR IN MHD DUCT = ,1X,F8.4   ,/,6X,
     5        10HPAUX/PSE =,1X,F8.4,/,6X,18H(P1-PCOMB)/PCOMB =,1X,F8.4,/,
     6         6X, 8HPC/MAW =,1X,1PE12.5, /,6X, 8HQA/MAW =,1X,E12.5, /,6X,
     7         7HTHETA =,4X,E12.5 ,/, 6X,6HPMHD =,4X,E12.5 , /, 6X,
     8         4HPE =,5X ,E12.5 )
 2800 FORMAT(1H0,1X,8HAIR SIDE ,8X,5HTOTAL ,3X,8HPRESSURE  ,3X,
     5         11HTEMPERATURE  ,
     9          //,4X,16HCOMPRESSOR INLET , 3X,F10.4, 4X,F10.4 , / ,
     A         4X,17HCOMPRESSOR OUTLET ,2X,F10.4, 4X,F10.4 , /,
     B         4X,18HAIR PREHEATER EXIT ,1X,F10.4, 4X,F10.4 )
 2900 FORMAT(1H0,1X,8HGAS SIDE   ,14X,5HTOTAL   ,3X,8HPRESSURE   ,3X,
     8         11HTEMPERATURE  ,
     C          //, 4X,14HMHD DUCT INLET,12X,F10.4,4X,F10.4 , / , 4X,
     D         13HDIFFUSER EXIT ,13X,F10.4 ,4X,F10.4 , /, 4X,
     E         13HINJECTOR EXIT ,13X,F10.4 ,4X,F10.4 , /, 4X,
     F         18HAIR PREHEATER EXIT , 8X, F10.4 , 4X, F10.4 , /, 4X,
     G         25HBOTTOMING HEAT EXCH. EXIT ,1X  ,F10.4 ,4X,F10.4 )
 3000 FORMAT(1H0,4X,3HC = ,1X,F8.4,/,5X,4HDK = ,1X,F8.4,/,5X,4HDB = ,
     1         1X,F8.4 )
 9000 CONTINUE
      CONTINUE
      CALL EXIT
      END
```

PRT,SC PCF.DBLFFF

4. DQHDUCT

```
EI*MHDDUCT.DQHDUCT
      IMPLICIT DOUBLE PRECISION (A-H, O-Z)
      INTEGER HEAD
      EXTERNAL FUNCT2, PWRE,FUNCT3
      DOUBLE PRECISION NA, NC, KD, MU, N, MUB, KP, K, MG,
     1     LMBDA1, LMBDA2
      COMMON      NT, NP, HEAD(24),
     1     VT(50), VP(50), ZMU(50,50), ZW(50,50), ZH(50,50),
     2     ZSIGMA(50,50), ZS(50,50), ZQADD(50,50), SC(10,10), SB(10,10)
      DIMENSION VBW(50),VBS(50)
      DIMENSION TT(10),HH(10),RHORHO(10),SS(10),
     1     X(99),RD(99),P(99),T(99),SIGMA(99),H(99),RHO(99),MU(99),
     2     U(99),S(99),N(99),B(99),MUB(99),VK(99)
      DIMENSION   DWG(7), WS(7), PS(7), PSE(7), PO(7), ETA(7)
      DIMENSION PRWT(31),PRAT(31),PROXT(31)
      COMMON /COMT2/ HAT(31), HWT(31),VTHWT(31),RWA,RWAP1,H2,HOXT(31),
     1     VHH(50),RMOX,RMAW,RMRP,VBT(50),NBT
      COMMON /REPR/HSTACK,TSTACK,PP(10),HRE
      DIMENSION LTAB1(10),LTAB2(41)
      DIMENSION   PSETAG(7), PSEPAU(7), CETA(7), PSERHO(7), VETAS(7)
      DATA PRWT  /   .2209D0,  .5412D0,  1.1305D0,  2.120D0,  3.683D0,
     1     6.038D0,  9.469D0,  1.4327D1,  2.105D1,  3.019D1,  4.241D1,
     2     5.853D1,  7.953D1,  1.0667D2,  1.4135D2,  1.0536D2, 2.408D2,
     3     3.100D2,  3.960D2,  5.022D2,  6.324D2,  7.915D2,  9.848D2,
     4     1.2184D3, 1.4994D3, 1.8356D3, 2.237D3,  2.713D3,  3.276D3,
     5     3.940D3,  4.720D3 /
      DATA PRAT  /   .485D0, 1.059D0, 2.005D0,  3.446D0,  5.526D0,
     1     8.911D0,  1.2298D1, 1.7413D1, 2.401D1,  3.1694D0, 4.288D0,
     2     5.586D1,  7.173D1,  9.095D1,  1.1403D2, 1.4151D2, 1.740D2,
     3     2.121D2,  2.566D2,  3.081D2,  3.676D2,  4.357D2,  5.135D2,
     4     6.019D2,  7.020D2,  8.148D2,  9.414D2,  1.0834D3, 1.2417D3,
     5     1.4180D3, 1.6132D3   /
      DATA PROXT /   1.8305D0,  4.005D0, 7.629D0,  1.3244D1, 2.153D1,
     1     3.330D1,  4.955D1,   7.146D1, 1.0036D2, 1.3780D2, 1.8554D2,
     2     2.455D2,  3.200D2,   4.114D2, 5.224D2,  6.560D2,  8.151D2,
     3     1.0038D3, 1.2252D3,  1.4837D3, 1.7844D3, 2.1310D3, 2.530D3,
     4     2.985D3,  3.504D3,   4.092D3, 4.756D3,  5.503D3, 6.349D3,
     5     7.282D3,  8.330D3  /
      DATA LTAB1/3H1  ,3H2  ,3H3  ,3H4  ,3H5  ,3H6  ,3H7  ,3H8  ,
     1     3H9  ,3H10  /
     2     LTAB2 / 3H400, 3H401, 3H402, 3H403, 3H404, 3H405, 3H406,
     3             3H407, 3H408, 3H409, 3H410, 3H411, 3H412, 3H413,
     4             3H414, 3H415, 3H416, 3H417, 3H418, 3H419, 3H420,
     5             3H421,3H422,3H423,3H424,3H425,3H426,3H427,3H428,
     5             3H429,3H430,3H431,3H432,3H433,3H434,3H435,3H436,
     5             3H437,3H438,3H439,3H440 /,
     6     LAB5, LAB6 / 3H5  , 3H6  /, ZERO / 0.0D0 /
      IN = 5
      IOUT = 6
      NZDIM = 50
      TEMP = 2.222222222D2
      DO 10 I=1,31
         VTHWT(I) = TEMP
         TEMP = TEMP + 5.5555555555D1
   10 CONTINUE
      TOLBIS = 1.DE-3
      TAMB = 288.3
```

```
      PAMB = 1.
      RWA  = .00639
      ETAC = 0.9
      WA = 28.970
      TSTACK = 0.0
      DPIN = .02
      DPCO = 0.05
      DPAP = 0.03
      DPSG = 0.06
      DPPN = 0.06
      DPINJ = 0.0
      DB = 2.0
      K0 = 0.82
      B0 = 6.0
      C = 0.10
      LMBDA1 = 0.05
      LMBDA2 = 0.1
      ETAD = 0.8
      F = 0.005
      ETAG = 0.984
      ETAI = 0.985
      PAUXP1 = 0.015
      DHA = 0.0
      RWAP1 = RWA + 1.0
      KP = 101325.0
      R = 8314.69/KP
      RLH = 1.0 + LMBDA2
      CALL SETUP(PHI)
  101 READ(IN, 2100, END=9999)  PHIIN, PMHD, TCOMB, PCOMB, U0,RMRP,RMOX
      IF(PHI  .EQ.  PHIIN) GO TO 102
      WRITE(IOUT,3100)
 3100 FORMAT(// ,35H PHI AND PHIIN ARE NOT CONSISTENT    , //,
     1        38H CHECK IF PROPER DISK  FILE IS INPUT      ,//
     2       ,20H RUN TERMINATED ****        )
      STOP
  102 CONTINUE
      READ(IN, 2100) RMAW, RMC, RMS, RMWC, RMWS, RMCW, RMAP
      READ(IN, 2100) WC, NC, NA, HHVDAF, XS, TSTACK
     1           ,FTEMP,AFGR
      WRITE(IOUT, 2200) HEAD
      WRITE(IOUT, 2210) PHIIN, PMHD, TCOMB, PCOMB, U0,RMRP,RMOX
      WRITE(IOUT, 2220) RMAW, RMC, RMS, RMWC, RMWS, RMCW, RMAP
      WRITE(IOUT, 2230) WC, NC, NA, HHVDAF, XS, TSTACK
     1           ,FTEMP,AFGR
C     FTEMP IS FUEL GAS TEMP TO COMB. DEG K.
C     AFGR IS RATIO OF AIR TO PRODUCT FUEL GAS
      READ(IN, 2110) NBT
      READ(IN, 2121)  (VBT(I), I=1,NBT)
      READ(IN, 2121)  (VBW(I), I=1,NBT)
      READ(IN, 2121)  (VBH(I), I=1,NBT)
      READ(IN, 2121)  (VBS(I), I=1,NBT)
 2121 FORMAT()
      DK = .05-.0025*PMHD*1.0E-8
      IF(RMAP .GT. 1.E-30)DPINJ = .03
      HSTACK=SINTPA(VBT,VBH,TSTACK,NBT)
      RMGP = RMAP + 1.0
      RMA = RMAW / RWAP1
```

REPRODUCIBILITY OF THE
ORIGINAL PAGE IS POOR

```
C      THIS PROGRAM IS FOR LOW BTU GAS PRODUCED FROM ILL NO.6 COAL
       RMAG = AFGR * RMCW
       TGA = 672.2
       DHGS =(1200.0 - 27.0) * 2325.98
       QPPC = 0.0
       IF(AFGR .GT. 0.0)  QPPC = 458.0 * 2325.98
       FGCRG = 4.28
       PGAS = 14400. / 0.4536
       STCR = 0.552
       HHVCO = 12028.0 * 2325.98
       TT(4) = TCOMB
       PP(4) = PCOMB
       PP(1) = PAMB*(1.-DPIN)
       PP(10) = PAMB
       TT(10) = TSTACK
       HH(10) = HSTACK
       PP(3) = (1.0+DPCO) * PP(4)
       PP(2) = (1.0+DPAP) * PP(3)
       PP(9) = (1.0+DPSG) * PP(10)
       PP(8) = (1.0+DPPR) * PP(9)
       PP(7) = (1.0+DPINJ) * PP(8)
       PE = PMHD / ETAI
       THETA = 2325.98 * HHVDAF
       DRP31 = (PP(2)-PP(4)) / PP(4)
C      ***           T1 CALCULATION
       T2=TAMB
C      H IS ENTHALPY OF MIXTURE PER POUND OF MOIST AIR
       HAMB =(SINTPA(VTHWT,HAT,T2,31)+RWA*SINTPA(VTHWT,HWT,T2,31))/RWAP1
C      PRWT,PRAT AND PROX ARE THE RELATIVE PRESS. FOR WATER VAPOR, AIR
C      AND OXYGEN TAKEN FROM GAS TABLES STARTING AT 400R(222.22K) WITH
C      100R(55.556K) INCREMENTS
       HH(1) = HAMB+SINTPA(VTHWT,HOXT,T2,31)*RMOX/RMAW+HSTACK/2325.98*
      1     RMRP / RMAW
       H2 = HH(1)
       TT(1) = FUNCT3(HH(1))
       PRA= SINTPA(VTHWT,PRAT,TAMB,31)*PP(2)/PP(1)
       PRW= SINTPA(VTHWT,PRWT,TAMB,31)*PP(2)/PP(1)
       DHAW=(SINTPA(PRAT,HAT,PRA,31)+RWA*SINTPA(PRWT,HWT,PRW,31))/RWAP1
      1     - HAMB
       PROX =SINTPA(VTHWT,PROXT,TAMB,31)*PP(2)/PP(1)
       DHOX=(SINTPA(PROXT,HOXT,PROX,31)- SINTPA(VTHWT,HOXT,TAMB,31))
       PCMAWR=(DHAW+RMOX/RMAW*DHOX)*2325.98/ETAC+RMRP/RMAW*PWRE(PP(10))
       PCMAWR = PCMAWR * (1.0 + RMAG / RMAW)
       HH(2) = HH(1)+PCMAWR/2325.98
       HH(2)=HH(2)/(1.0+RMAG/RMAW)
       H2 = HH(2)
       TT(2)=FUNCT3(HH(2))
       FSH = 0.0
       IF(FTEMP .LT. 300.)  GO TO 113
C      GTEMP IS TEMP OF GAS DELIVERED BY GASIFIER DEG.F.
       GTEMP = 1144.4
       FSH = SEHE(FTEMP)
C      ACCOUNTS FOR SENSIBLE HEAT ADDED TO FUEL GAS, QFG IS JOULES/KG
C      /MOLE
       FSG = SEHE(GTEMP)
       QFG = FSH - FSG
   113 CONTINUE
```

```
C       ***          T2 CALCULATION
              QADD = ZTP(ZQADD, TT(4), PP(4), NZDIM, NT, NP, VT, VP)
        QADD = QADD - FSH
        T2 = 298.16
C       H IS ENTHALPY OF MIXTURE PER POUND OF MOIST AIR
        HCOMB=ZTP(ZH,TT(4),PP(4),NZDIM,NT,NP,VT,VP)
        HAP = (SINTPA(VTHWT,HAT,TSTACK,31)+RWA*SINTPA(VTHWT,HWT,TSTACK,31)
       1      ) / RWAP1
        HAO = (SINTPA(VTHWT,HAT,T2,31)+ RWA* SINTPA(VTHWT,HWT,T2,31))/
       1      RWAP1
        QADDP = 0.4536*WC*WC*(QADD/WC/WC+LMBDA1*THETA)/1055.0/RWAP1/NA/
       1      XS/PHI/WA
        HA = HAO + QADDP
        HH(3) = HAO+(1.0+RMRP*RMAP/RMAW)*QADDP-RMRP*RMAP/RMAW*(HA-HAP)+
       1      RMRP/RMAW*(1.0-RMAP)*HCOMB/2325.98
        H2 = HH(3)
         TT(3)=FUNCT3(H2)
        HGA= (SINTPA(VTHWT,HAT,TGA,31)+RWA*SINTPA(VTHWT,HWT,TGA,31))/RWAP1
        QAMAWR = (H2-HH(2))*2325.98+(HGA-HH(2))*RMAG/RMAW*2325.98
              HH(4) = ZTP(ZH, TT(4), PP(4), NZDIM, NT, NP, VT, VP)
              SS(4) = ZTP(ZS, TT(4), PP(4), NZDIM, NT, NP, VT, VP)
              RHORHO(4) = ZTP(ZW, TT(4), PP(4), NZDIM, NT, NP, VT, VP) *
       1                  PP(4)/(R*TT(4))
        WRITE(IOUT,2500)ETAD,ETAG,ETAI
        WRITE(IOUT,2600)LMBDA1,LMBDA2
        WRITE(IOUT,2700)F,PAUXP1,DRP31,PCMAWR,QAMAWR,THETA,PMHD,PE
       1      ,WFG
        WRITE(IOUT,3000)C,DK,DB
              RO(1) = 1.0
              X(1) = 0.0
              S(1) = SS(4)
              SS(5) = SS(4)
        U(1) = U0
              HH(5) = HH(4) - 0.5*U(1)*U(1)
              H(1) = HH(5)
              TT(5) = TZZ(S(1), ZS, H(1), ZH, NZDIM, NT, NP, VT, VP)
              T(1) = TT(5)
              P6T = PZT(S(1), ZS, T(1), NZDIM, NT, NP, VT, VP)
              PP(5) = P6T
              P(1) = P6T
              SIGMA(1) = ZTP(ZSIGMA, T(1), P(1), NZDIM, NT, NP, VT, VP)
              RHORHO(5) = ZTP(ZW, T(1), P(1), NZDIM, NT, NP, VT, VP) *
                          P(1)/(R*T(1))
              RHO(1) = RHORHO(5)
              MU(1) = ZTP(ZMU, T(1), P(1), NZDIM, NT, NP, VT, VP)
        K = K0
        VK(1) = K
        B(1) = B0
              MUB(1) = B(1) * MU(1)
              JSAFE = 5
        PSS = PP(7) * 1.015
              MG = PE / ((ZPZ(ZH, PSS, SS(4), ZS, NZDIM, NT, NP, VT, VP)
       1                - HH(4)) * (0.05 - K))
        IF(MG .LE. 0.0)  MG = 1300.0 * PMHD / 1.3E9
              KSAFE = JSAFE
  201         CONTINUE
                    DN = 0.0
```

```
                  IY = 1
                  IQ = 0
                  D4 = DSQRT(MG/(U(1)*RHO(1)))
                  RD4 = 1.0/D4
                  FXI = RHO(1)*U(1)*F*2.0/(B(1)**2*SIGMA(1)*D4)
                  N(1) = K*(1.0-K)/(1.0-K+FXI)
                  IV = 10
                  DP = 10.0
                  DP6T = 10.0
                  P6T = 10.0
                  JCNT = 11
C     ***      REPLACE XP CALL , II = XP(P(1), 0) + 2
            DO 290 I=1,NP
            IF (P(1) .GE. VP(I))    GO TO 290
                  II = I
                  GO TO 295
      290      CONTINUE
            II = NP
      295   CONTINUE
C                       P6T LOOP
      301   CONTINUE
                  IF (MOD(IQ,2) .EQ. 0)  GO TO 305
                        P6T = P6T + DP6T
                        P(IY) = (P6T-PP(7))/DP6T*DP + P(IY-1)
                        DP = P(IY-1) - P(IY-2)
                        DN = N(IY-1) - N(IY-2)
                        FXI = FXIM - FXI
                        GO TO 315
      305         CONTINUE
                        IY = IY+1
                        II = II-1
C     ***   REPLACE PX CALL ,  P(IY) = PX(II, 0)
                  P(IY) = VP(II)
      315         HS = ZPZ(ZH, P(IY), S(IY-1), ZS, NZDIM, NT, NP, VT,
     1                  VP)
                  DHS = HS - H(IY-1)
                  TEMP = 2.0*C*DHS + U(IY-1)**2
                  IF (TEMP .GE. 0.000)    GO TO 325
                        WRITE(IOUT, 2310)  TEMP, C
                  GO TO 500
      325         U(IY) = DSQRT(TEMP)
                  DPOLD = DP
                  DP = P(IY) - P(IY-1)
                  N(IY) = DN/DPOLD*DP + N(IY-1)
                  FXIM = 2.0*FXI
C                       VARY FXIM, CONVERGENCE BASED ON N
                  FACTX = 8.0*KP*(1.0+C)
                  FACTN = 0.5*RLH*(1.0+C)
                  FT2 = 2.0*F
C
      351   CONTINUE
                  H(IY) =  (FACTN*(N(IY)+N(IY-1))-C)*DHS + H(IY-1)
                  T(IY) = TPZ(P(IY), H(IY), ZH, NZDIM, NT, NP, VT, VP)
                  RHO(IY) = ZTP(ZH, T(IY), P(IY), NZDIM, NT, NP, VT, VP)
     1                  *P(IY)/(R*T(IY))
                  SIGMA(IY) = ZTP(ZSIGMA, T(IY), P(IY), NZDIM, NT, NP,
     1                  VT, VP)
```

```
                            RD(IY) = DSQRT(MG/(U(IY)*RHO(IY))) * RD4
                            K = VK(IY-1) + DK*DHS*RHO(IY)/101300.
                            VK(IY) = K
                            B(IY) = B(IY-1)
                            IF (P(IY) .LE. 2.0)  B(IY) = B(IY) +
               1                 DB*DHS*RHO(IY)/101300.0
                            FXIOLD = FXI
                            FXI = FT2*RHO(IY)*U(IY) / (SIGMA(IY)*RD(IY)*D4*
               1                      B(IY)**2)
                            FXIM = FXIM - FXIOLD + FXI
                            OLDNY = N(IY)
                            N(IY) = K*(1.0-K)/(1.0-K+FXI)
                            IF (DABS(OLDNY-N(IY)) .LT. 1.0E-5)     GO TO 355
                            JCNT = JCNT -1
                            IF (JCNT .LE. 1)  GO TO 355
                      GO TO 351
            355             X(IY) = DP / ((K-0.5*FXIM-1.0) * (SIGMA(IY-1)+SIGMA(IY
               1                 )) * (B(IY-1)**2 + B(IY)**2) * (U(IY-1)+U(IY))) *
               2                 FACTX + X(IY-1)
                            S(IY) = ZTP(ZS, T(IY), P(IY), NZDIM, NT, NP, VT, VP)
                            MU(IY) = ZTP(ZMU, T(IY), P(IY), NZDIM, NT, NP, VT, VP)
                            MUB(IY) = MU(IY)*B(IY)
                            DN = N(IY) - N(IY-1)
                            JCNT = 4
                            IF (P(IY) .GT. PP(7))     GO TO 365
                                 HH(7) = 0.5*U(IY)**2 + H(IY)
                                 TSS = TZZ(S(IY), ZS, HH(7), ZH, NZDIM, NT, NP,
               1                           VT, VP)
                                 PSS = PZT(S(IY), ZS, TSS, NZDIM, NT, NP, VT, VP)
                                 P6TOLD = P6T
                                 P6T = ETAD*(PSS-P(IY)) + P(IY)
                                 DP6T = P6TOLD - P6T
            365             Q = IQ
                            IQ = 0
                            IF ((P6T-10.D*Q) .LE. PP(7))  IQ = 1
         C                       CONVERGENCE TEST FOR P6T
                            IV = IV-IQ
                            IF (IV .LT. 0)  GO TO 375
                            IF (DABS(DP6T) .LT. 1.0D-10)  GO TO 375
                            IF (DABS(PP(7)-P6T) .LT. 0.005)  GO TO 375
                      GO TO 301
            375 CONTINUE
                            WRITE(IOUT, 2300)  MG
                            OLDMG = MG
                            MG = PE*RLH/(HH(4)-HH(7))
                            IF (DABS(OLDMG/MG-1.0) .LT. 2.0D-4)  GO TO 405
                            KSAFE = KSAFE - 1
                            IF (KSAFE .LE. 0)  GO TO 405
                      GO TO 201
         C                       END OF KSAFE (FXI) LOOP
            405 CONTINUE
                      HH(6) = H(IY)
                      SS(6) = S(IY)
                      PP(6) = P(IY)
                      TT(6) = T(IY)
                      RHORHO(6) =  RHO(IY)
                      H5SP = ZPZ(ZH, PSS, SS(4), ZS, NZDIM, NT, NP, VT, VP)
```

```
              ETAMHD = PE / (MG*(HH(4)-H5SP))
              TT(7) = TPZ(PP(7), HH(7), ZH, NZDIM, NT, NP, VT, VP)
              SS(7) = ZTP(ZS, TT(7), PP(7), NZDIM, NT, NP, VT, VP)
              RHORHO(7) = ZTP(ZW, TT(7), PP(7), NZDIM, NT, NP, VT, VP)
     1                          PP(7)/(R*TT(7))
              TT(8) = TT(7)
              SS(8) = SS(7)
              RHORHO(8) = RHORHO(7)
              HH(8) = HH(7)
              IF (RHAP .LE. 1.0D-30)  GO TO 425
  411              DHA =(SINTPA(VTHWT, HAT, TT(8), 31) +
     1                 SINTPA(  VTHWT,HWT,TT(8),31)*RWA)/RWAP1-HAMB
                   HH(8) = (RHAP*DHA + HH(7)) / (RHAP+1.0)
                   TT8NEW = SINTPA(V8H, V8T, HH(8), N8T)
                   IF (DABS(TT(8)-TT8NEW) .LE. 0.1) GO TO 415
                         TT(8) = TT8NEW
              GO TO 411
  415              TT(8) = TT8NEW
                   SS(8) = SINTPA(V8T, V8S, TT(8), N8T)
                   RHORHO(8) = SINTPA(V8T, V8W, TT(8), N8T) * PP(8)/
     1                         (R*TT(8))
  425          HH(9) = HH(8)-DAHAWR/RMGP*RMAW-GFG/WC*RMC/RMGP
              TT(9) = SINTPA(V8H, V8T, HH(9), N8T)
              SS(9) = SINTPA(V8T, V8S, TT(9), N8T)
              RHORHO(9) = SINTPA(V8T, V8W, TT(9), N8T) * PP(9)/(R*TT(9))
              DWI = PE - PMHD
         QS0 = MG*RMGP*(HH(9)-HH(10))-(QPPC-STCR*DHGS)*RMCW/FGCRG* MG
         QC = RMC * HHVCU * MG / FGCRG
              QS1 = LMBDA1 * QC
              QS2 = LMBDA2 * PE
              PC = PCMAWR * RMAW * MG
              ETAS = 0.39
        RETAG = 1.0 / ETAG
              DO 480  KY=1,7
                   ETAS = ETAS + 0.01
                   ETASF = 0.995 * ETAS
                   VETAS(KY) = ETAS
                   PS(KY) = ((1.0-ETAG)*PC -QS0 - QS1 - QS2 - DWI)*ETASF/
     1                   ((RETAG-1.0)*ETASF-1.0)
                   PSE(KY) = PS(KY) - PC*ETAG
                   DWG(KY) = PSE(KY) * RETAG - PSE(KY)
                   QS(KY) = 0.995 * (DWG(KY)+QS0+QS1+QS2+DWI)
                   PO(KY) = (1.0-PAUXP1)*PSE(KY) + PMHD
     1             - PGAS * RMC * MG / FGCRG
                   ETA(KY) = PO(KY)/QC
  480         CONTINUE
C                  OUTPUT
              LPP3 = 10.0*PP(4)
              LK = 100.0*K
              YY = 10.0*DRP31
              IV = YY + 0.5
              KY = 10.0*(YY-IV)
              LU0 = U(1)
              LC = 100.0*C + 0.5
              LTT3 = TT(4)
              LB0 = B(1)
              RMRPMG = RMRP*MG
```

```
                  RMOXMG = RMOX*MG
                  RMAWMG =  RMAW *MG
            RMAMG = RMA*MG
                  RMGPMG=  RMGP *MG
                  RMAPMG = RMAP *MG
                  RMCMG  = RMC *MG
                  RMCWMG = RMCW *MG
                  RMSMG  = RMS* MG
                  RMWCMG = RMWC *MG
                  RMWSMG = RMWS *MG
                  RMARP = RMAWMG+RMAPMG
                  RGMAMG = MG +RMAPMG
            WRITE(IOUT,2800) PP(1), TT(1),PP(2),TT(2),PP(3),TT(3)
            WRITE(IOUT,2900) PP(4),TT(4), PP(7),TT(7),PP(8),TT(8),
     1            PP(9),TT(9), PP(10),TT(10)
                  WRITE(IOUT, 2420)
            WRITE(IOUT,2424)PAMB,TAMB,HAMB,RMAWMG
            WRITE(IOUT,2423) LTAB1(1),PP(1),TT(1),HH(1),RMAWMG
            WRITE(IOUT,2423) LTAB1(2),PP(2),TT(2),HH(2),RMARP
            WRITE(IOUT,2423) LTAB1(3),PP(3),TT(3),HH(3),RMARP
            WRITE(IOUT,2425)(LTAB1(J),PP(J),TT(J),HH(J),RHORHO(J),SS(J),MG,
     1            J=4,6 )
            WRITE(IOUT,2425)LTAB1(7),PP(7),TT(7),HH(7),RHORHO(7),SS(7),MG
            WRITE(IOUT,2425)(LTAB1(J),PP(J),TT(J),HH(J),RHORHO(J),SS(J),
     1            RGMAMG,J=8,10)
                  WRITE(IOUT, 2430)
                  JY = IY - 1
                  IF (JY .LE. 0)     GO TO 485
                  WRITE(IOUT,2435)(LTAB2(J),X(J), RD(J), P(J), T(J), SIGMA(J),
     1            H(J), RHO(J), U(J), MU(J), S(J), N(J), MUB(J),
     2            VK(J), B(J), J=1,JY)
  485             J = IY
                  WRITE(IOUT, 2435)  LAB5, X(J),RD(J), P(J), T(J), SIGMA(J),
     1                  H(J), RHO(J), U(J), MU(J), S(J), N(J), MUB(J)
     2            , VK(J), B(J)
                  V = 0.5 * D4 * (RD(IY)+1.0)
                  XVD4 = (X(IY) + V) * V
                  D5 = RD(IY) * D4
                  A4 = D4 * D4
                  A5 = D5*D5
                  WRITE(IOUT, 2440)  MG, RMAMG, RMAWMG, RMGPMG, RMAPMG,
     1                  RMCMG, RMCWMG, RMSMG, RMWCMG, RMWSMG, XVD4, D4, D5,
     2                  A4, A5, WC, ETAMHD
                  DO 490 KY=1,7
                        PSETAG(KY) = PSE(KY)/ETAG
                        PSEMHD(KY) = PSE(KY) + PMHD
                        PSEPAU(KY) = PSE(KY) * PAUXPL
                        CETA(KY) = 3412.16/ETA(KY)
  490             CONTINUE
            WRITE(IOUT,2460)
            WRITE(IOUT,2480)(VETAS(J),J=1,7)
            WRITE(IOUT,2470)(QS0,J=1,7),(QS1,J=1,7)
     1                  , (QS2,J=1,7), (DQ1,J=1,7), (DQG(J),J=1,7),
     2                  (QS(J),J=1,7), (PS(J),J=1,7),
     3                  (PSETAG(J),J=1,7), (PC,J=1,7), (PSE(J),J=1,7),
     4                  (PMHD,J=1,7), (PSEMHD(J), J=1,7), (PSEPAU(J),J=1,7),
     5            (PO(J),J=1,7),(CETA(J),J=1,7)
```

```
      WRITE(IOUT,2490)(ETA(J),J=1,7)
  499 CONTINUE
  500 CONTINUE
      GO TO 101
 2100 FORMAT(8F10.0)
 2110 FORMAT(I2)
 2120 FORMAT(6F10.0)
 2200 FORMAT(1H1, 12A6 / 1H , 12A6)
 2210 FORMAT(1H0, 6HPHI = , F6.2, 5X, 6HPMHD =, 1PE12.5, 5X, 7HTCOMB =,
     1       0PF8.1, 5X, 7HPCOMB =, F8.3, 5X, 4HU0 = , F8.2,
     2       /,1X,7HRMNP = ,F10.3,5X,7HRMOX = ,F10.3)
 2220 FORMAT(1H0, 6HRMAW =, 1PE14.6, 5X, 6H RMC =,E14.6, 5X,  6H RMS =,
     1       E14.6 /1H , 6HRMWC =, E14.6, 5X, 6HRMWS =, E14.6, 5X,
     2       6HRMCW =, E14.6, 5X, 6HRMAP =, E14.6)
 2230 FORMAT(1H0, 4HNC = , F8.2, 5X, 4HNC =, F8.4, 5X, 4HNA = F8.4,
     1       5X,8HHHVDAF =, 1PE14.6,4HXS =,E14.6,/1H,
     2       8HTSTACK =,E14.6,5X,10HTFUELGAS =,E14.6,5X,8HAIR/FG =,E14.6)
 2240 FORMAT(1H0, 24H**ITERATION ON T1 FAILED)
 2250 FORMAT(1H0, 24H**ITERATION ON T2 FAILED)
 2300 FORMAT(1H , 6H     MG, 1PE13.6)
 2310 FORMAT(1H0/1H0, 42H** CALCULATION OF FLUID VELOCITY INVOLVES ,
     1       22HTAKING SQUARE ROOT OF , 1PE12.5, 1H. /1H , 5X,
     2       21HCURRENT VALUE OF C = , E12.5,
     3       43H. HIGHER VALUES OF C WILL ALSO BE IGNORED. )
 2400 FORMAT(1H , 2X, 2(1H-, I2), 1H-, 2I1, 1H-, I3, 1H-, I2, 1H-, I4,
     1       1H-, I1)
 2410 FORMAT(1H0, 3X, 5HINPUT/1H0, 10X, 5H PHI , F5.2, 7H CYCLE , I2,
     1       4H P3 , F6.2, 4H T3 , F5.0,12H (P1-P3)/P3 , F6.2,
     2       3H U , F4.0, 3H C , F5.2, 9H LAMBDA1 , F5.2, 9H LAMDA2  ,
     3       F5.2 /1H0, 3H K , F5.2, 4H PE , 1PE8.1, 4H P6 , 0PF6.2,
     4       6H ETAD , F5.2, 3H F , F6.3, 4H T8 , F4.0, 3H B , F2.0,
     5       8H PC/MAW , F7.0, 4H P1 , F6.2, 4H T1 , F5.0, 4H P2 , F6.2
     6       /1H0, 4H T2 , F5.0, 6H ETAG , F6.3, 6H ETAI , F6.3,
     7       10H PAUX/PSE , F6.3, 8H QA/MAW , F7.0, 7H THETA , 1PE10.3
     8       /     , 6H ETAS , 0PF6.2, 6(1H, F6.2)))
 2420 FORMAT([illegible] .//,11X,        10HGAS STATES  .//,2X,6HPOINT,5X,1HP,
     1       [illegible]T, 10X, 1HH, 10X, 3HRHO, 9X, 1HS,9X,1HM,//)
 2425 FO[illegible] 2X, A3, 4X, F6.3, 4X, F5.0, 4X, F[illegible], 4X, F7.4,
     1               4X,F8.3)
 2430 FORM[illegible]
     1       [illegible]
     2       /X, [illegible] /X, [illegible]
     3       6X, [illegible] 5X, 1HK, 5X, [illegible]
 2435 FORMAT(1H , 3X, A3, 2X, F7.3, 2X, F6.3, 2X, F6.3, 2X, F5.0,
     1       2X,F6.3,2X,F9.0,2X,F7.4,2X,F6.1,2X,F7.4,2X,F7.1,
     2       2X, F6.4, 2X, F7.4, F6.3, F6.2)
 2440 FORMAT(1H0, 3X, 25HMASSFLOWS, GENERATOR AREA
     1       /1H , 5X, 4H MG , F8.3, 4H MA , F8.3, 4H MAW, F8.3,
     2             4H MG*, F8.3, 4H MA*, F8.3
     3       /1H , 5X, 4H MC , F8.3, 4H MCW, F8.3, 4H MS , F8.3,
     4             4H MWC, F8.3, 4H MWS, F8.3
     5       /1H , 6X, 15H (X + DM) * DM , F9.5, 3H D4, F7.4,
     6             3H D5, F7.4, 3H A4, F7.4, 3H A5, F7.4
     7       /1H , 5X, 12H THETA * MC , 1PE12.5, 8H ETAMHD , 0PF8.5)
 2423 FORMAT(1H ,2X,A3,4X,F6.3,4X,F5.0,4X,F9.0,22X,4X,F8.3)
 2424 FORMAT(1H ,2X,3H0  ,4X,F6.3,4X,F5.0,4X,F9.0,22X,4X,F8.3)
 2460 FORMAT(1H0,13X,         19HOVERALL PERFORMANCE  )
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      2470 FORMAT(1HH,4X,8HQS0          ,3X,7E14.5
          3       /1H , 4X, 8HQS1          ,3X,7E14.5
          4       /1H , 4X, 8HQS2          ,3X,7E14.5
          5       /1H , 4X, 8HDQI          ,3X,7E14.5
          6       /1H , 4X, 8HDQG          ,3X,7E14.5
          7       /1H , 4X, 8HQS           ,3X,7E14.5
          8       /1H , 4X, 8HPS           ,3X,7E14.5
          9       /1H , 4X, 8HPT           ,3X,7E14.5
          1       /1H , 4X, 8HPC           ,3X,7E14.5
          2       /1H , 4X, 8HPSE          ,3X,7E14.5
          3       /1H , 4X, 8HPMHD         ,3X,7E14.5
          4       /1H , 4X, 8HPSE+PMHD,3X,7E14.5
          5       /1H , 4X, 8HPAUX         ,3X,7E14.5
          6       /1H , 4X, 8HP            ,3X,7E14.5
          7       /1H , 4X, 8HHR           ,3X,7E14.5)
      2480 FORMAT(1H0    ,11HSTEAM PLANT   ,/ ,1X,10HEFFICIENCY ,7(2X,F12.2))
      2490 FORMAT(1H0    ,7HOVERALL ,/,1X,11HEFFICIENCY   ,2X,0PF12.4,
          1       6(2X,F12.4))
      2500 FORMAT(1H0,1X,12HEFFICIENCIES  /,5X,8HDIFFUSER ,12X,F10.4 /
          1       ,5X,18HROTATING GENERATOR ,2X,F10.4 , /
          2       ,5X,14HDC/AC INVERTER ,6X,F10.4   )
      2600 FORMAT(1H0,1X,19HHEAT TRANSFER RATIO , / , 6X,
          3       37HFROM COMBUSTER TO SUBPOSED PLANT     = , F8.3, / ,6X,
          4       37HMHD GENERATOR TO SUBPOSED PLANT      = , F8.3 )
      2700 FORMAT(//,6X,29HFRICTION FACTOR IN MHD DUCT = ,1X,F8.4    ,/,6X,
          5     10HPAUX/PSE =,1X,F8.4,/,6X,18H(P1-PCOMB)/PCOMB =,1X,F8.4,/,
          6       6X, 8HPC/MAW =,1X,1PE12.5, /,6X, 8HQA/MAW =,1X,E12.5, /,6X,
          7       7HTHETA =,4X,E12.5 ,/, 6X,6HPMHD =,4X,E12.5 , /, 6X,
          8       4HPE =,5X,E12.5,/,6X,5HQFG =,E12.5)
      2800 FORMAT(1H0,1X,8HAIR SIDE ,8X,5HTOTAL ,3X,8HPRESSURE   ,3X,
          5       11HTEMPERATURE   ,
          9        //,4X,16HCOMPRESSOR INLET , 3X,F10.4, 4X,F10.4 , / ,
          A       4X,17HCOMPRESSOR OUTLET ,2X,F10.4, 4X,F10.4 , /,
          B       4X,18HAIR PREHEATER EXIT ,1X,F10.4, 4X,F10.4 )
      2900 FORMAT(1H0,1X,8HGAS SIDE   ,14X,5HTOTAL   ,3X,8HPRESSURE   ,3X,
          8       11HTEMPERATURE    ,
          C        //, 4X,14HMHD DUCT INLET,12X,F10.4,4X,F10.4 , / , 4X,
          D       13HDIFFUSER EXIT ,13X,F10.4  ,4X,F10.4 , /, 4X,
          E       13HINJECTOR EXIT ,13X,F10.4  ,4X,F10.4 , /, 4X,
          F       18HAIR PREHEATER EXIT , 8X, F10.4 , 4X, F10.4 , /, 4X,
          G       25HBOTTOMING HEAT EXCH. EXIT ,1X  ,F10.4 ,4X,F10.4 )
      3000 FORMAT(1H0,4X,3HC = ,1X,F8.4,/,5X,4HDK = ,1X,F8.4,/,5X,4HDB = ,
          I       1X,F8.4 )
      9999 CALL EXIT
           END
```

PRT,SC   PCF.DUCABS

5. FRECIRCINJOX

```
•MHDDUCT.FRECIRCINJOX
              IMPLICIT DOUBLE PRECISION (A-H, O-Z)
              INTEGER HEAD
              EXTERNAL FUNCT2,  PWRE,FUNCT3
              DOUBLE PRECISION NA, NC, KO, MU, N, MUB, KP, K, MG,
             1       LMBDA1, LMBDA2
              COMMON        NT, NP, HEAD(24),
             1       VT(50), VP(50), ZMU(50,50), ZW(50,50), ZH(50,50),
             2       ZSIGMA(50,50), ZS(50,50), ZQADD(50,50), SC(10,10), SB(10,10)
              DIMENSION VBW(50),VBS(50)
              DIMENSION TT(10),HH(10),RHORHO(10),SS(10),
             1              X(24), RD(24), P(24), T(24), SIGMA(24), H(24),
             2              RHO(24), MU(24), U(24), S(24), N(24), B(24),
             3              MUB(24), VK(24)
              DIMENSION   DWG(7), WS(7), PS(7), PSE(7), PO(7), ETA(7)
              DIMENSION PRAT(31),PRAT(31),PROXT(31)
              COMMON /COMT2/ HAT(31), HWT(31),VTHWT(31),RWA,RWAP1,H2,HOXT(31),
             1         VBH(50),RMOX,RMAW,RMRP,VBT(50),NBT
              COMMON /REPR/HSTACK,TSTACK,PP(10),HRE
              DIMENSION LTAB1(10), LTAB2(24)
              DIMENSION    PSETAG(7), PSEPAU(7), CETA(7), PSEMHD(7), VETAS(7)
              DATA  PRAT  /   .2209D0,  .5412D0,  1.1305D0,  2.120D0,  3.683D0,
             1      6.038D0,  9.469D0,  1.4327D1,  2.105D1,   3.019D1,  4.241D1,
             2      5.853D1,  7.953D1,  1.0667D2,  1.4135D2,  1.8536D2, 2.408D2,
             3      3.100D2,  3.960D2,  5.022D2,  6.324D2,   7.915D2,  9.848D2,
             4      1.2184D3, 1.4994D3, 1.8356D3, 2.237D3,   2.713D3,  3.276D3,
             5      3.940D3,  4.720D3 /
              DATA  PRAT  /   .4858D0, 1.0590D0, 2.005D0,  3.446D0,  5.526D0,
             1      8.411D0,  1.2298D1, 1.7413D1, 2.4010D1,  3.1694D0, 4.288D0,
             2      5.586D1,  7.173D1,  9.095D1,  1.1403D2, 1.4151D2, 1.740D2,
             3      2.121D2,  2.566D2,  3.081D2,  3.676D2,  4.357D2,  5.135D2,
             4      6.019D2,  7.020D2,  8.148D2,  9.414D2,  1.0834D3, 1.2417D3,
             5      1.4180D3, 1.6132D3  /
              DATA  PROXT /   1.8305D0,   4.005D0, 7.629D0,   1.3244D1, 2.153D1,
             1      3.330D1,  4.955D1,   7.146D1, 1.0036D2, 1.3760D2, 1.8554D2,
             2      2.455D2,  3.200D2,   4.114D2, 5.224D2,  6.560D2,  8.151D2,
             3      1.0038D3, 1.2252D3,  1.4839D3, 1.7844D3, 2.131D3, 2.530D3,
             4      2.985D3,  3.504D3,  4.092D3,  4.756D3,  5.503D3, 6.344D3,
             5      7.282D3,  8.330D3   /
              DATA LTAB1/3H1   ,3H2   ,3H3   ,3H4   ,3H5   ,3H6   ,3H7   ,3H8   ,
             1         3H9   ,3H10   /
             2         LTAB2 / 3H400, 3H401, 3H402, 3H403, 3H404, 3H405, 3H406,
             3                 3H407, 3H408, 3H409, 3H410, 3H411, 3H412, 3H413,
             4                 3H414, 3H415, 3H416, 3H417, 3H418, 3H419, 3H420,
             5                 3H421, 3H422, 3H423 /,
             6         LAB5, LAB6 / 3H5   , 3H6    /,    ZERO / 0.0D0 /
              IN = 5
              IOUT = 6
              NZDIM = 50
              TEMP = 2.2222222222D2
              DO 10 I=1,31
                VTHWT(I) = TEMP
                TEMP = TEMP + 5.5555555555D1
           10 CONTINUE
              TOLBIS = 1.0E-3
              TAMB = 288.3
              PAMB = 1.
```

```
      RWA  = .00639
      ETAC = 0.9
      WA = 28.970
      TSTACK = 0.0
      DPIN = .02
      DPCO = 0.05
      DPAP = 0.03
      DPSG = 0.06
      DPPH = 0.06
      DPINJ = 0.0
      DB = 2.0
      KO = 0.82
      BO = 6.0
      C = 0.10
      LMBDA1 = 0.05
      LMBDA2 = 0.1
      ETAD = 0.8
      F = 0.005
      ETAG = 0.984
      ETAI = 0.985
      PAUXP1 = 0.015
      DHA = 0.0
      RWAP1 = RWA + 1.0
      KP = 101325.0
      R = 8314.69/KP
      RLH = 1.0 + LMBDA2
      CALL SETUP(PHI)
  101 READ(IN, 2100, END=9999)  PHIIN, PMHD, TCOMB, PCOMB, UO,RMRP,RMOX
      IF(PHI  .EQ.  PHIIN) GO TO 102
      WRITE(IOUT,3100)
 3100 FORMAT(// ,35H PHI AND PHIIN ARE NOT CONSISTENT    , //,
     1          38H CHECK IF PROPER DISK  FILE IS INPUT     ,//
     2          ,20H RUN TERMINATED ****      )
      STOP
  102 CONTINUE
      READ(IN, 2100) RMAW, RMC, RMS, RMWC, RMWS, RMCW, RMAP
      READ(IN, 2100) WC, NC, NA, HHVDAF, XS, TSTACK
      WRITE(IOUT, 2200) HEAD
      WRITE(IOUT, 2210) PHIIN, PMHD, TCOMB, PCOMB, UO,RMRP,RMOX
      WRITE(IOUT, 2220) RMAW, RMC, RMS, RMWC, RMWS, RMCW, RMAP
      WRITE(IOUT, 2230) WC, NC, NA, HHVDAF, XS, TSTACK
      READ(IN, 2110) NBT
      READ(IN, 2121)  (VBT(I), I=1,NBT)
      READ(IN, 2121)  (VBW(I), I=1,NBT)
      READ(IN, 2121)  (VBH(I), I=1,NBT)
      READ(IN, 2121)  (VBS(I), I=1,NBT)
 2121 FORMAT()
      DK = .05-.0025*PMHD*1.0E-8
      IF(RMAP .GT. 1.E-30)DPINJ = .03
      HSTACK=SINTPA(VBT,VBH,TSTACK,NBT)
      RMGP = RMAP + 1.0
      RMA = RMAW / RWAP1
      TT(4) = TCOMB
      PP(4) = PCOMB
      PP(1) = PAMB*(1.-DPIN)
      PP(10) = PAMB
      TT(10) = TSTACK
```

```
      HH(10) = HSTACK
      PP(3) = (1.0+DPCO) * PP(4)
      PP(2) = (1.0+DPAP) * PP(3)
      PP(9) = (1.0+DPSG) * PP(10)
      PP(8) = (1.0+DPPR) * PP(9)
      PP(7) = (1.0+DPINJ) * PP(8)
      PE = PMHD / ETAI
      THETA = 2325.98 * HHVDAF
      DRP31 = (PP(2)-PP(4)) / PP(4)
C     ***          T1 CALCULATION
      T2=TAMB
C     H IS ENTHALPY OF MIXTURE PER POUND OF MOIST AIR
      HAMB =(SINTPA(VTHWT,HAT,T2,31)+RWA*SINTPA(VTHWT,HWT,T2,31))/RWAP1
C     PRWT,PRAT AND PROX ARE THE RELATIVE PRESS. FOR WATER VAPOR, AIR
C     AND OXYGEN TAKEN FROM GAS TABLES STARTING AT 400R(222.22K) WITH
C     100R(55.556K) INCREMENTS
      PRA= SINTPA(VTHWT,PRAT,TAMB,31)*PP(2)/PP(1)
      PRW= SINTPA(VTHWT,PRWT,TAMB,31)*PP(2)/PP(1)
      DHAW=(SINTPA(PRAT,HAT,PRA,31)+RWA*SINTPA(PRWT,HWT,PRW,31))/RWAP1
     1     - HAMB
      PROX =SINTPA(VTHWT,PROXT,TAMB,31)*PP(2)/PP(1)
      DHOX=(SINTPA(PROXT,HOXT,PROX,31)- SINTPA(VTHWT,HOXT,TAMB,31))
      PCMAWR=(DHAW+RMOX/RMAW*DHOX)*2325.98/ETAC+RMRP/RMAW*PARE(PP(10))
      HH(1)=HAMB+SINTPA(VTHWT,HOXT,T2,31)*RMOX/RMAW + HSTACK/2325.98*
     1     RMRP/RMAW
      H2 = HH(1)
      TT(1) = FUNCT3(HH(1))
      HH(2) = HH(1)+PCMAWR/2325.98
      H2 = HH(2)
      TT(2)=FUNCT3(HH(2))
C     ***          T2 CALCULATION
           QADD = ZTP(ZQADD, TT(4), PP(4), NZDIM, NT, NP, VT, VP)
     1
      T2 = 298.16
C     H IS ENTHALPY OF MIXTURE PER POUND OF MOIST AIR
      HCOMB=ZTP(ZH,TT(4),PP(4),NZDIM,NT,NP,VT,VP)
      HAP = (SINTPA(VTHWT,HAT,TSTACK,31)+RWA*SINTPA(VTHWT,HWT,TSTACK,31)
     1     ) / RWAP1
      HAO = (SINTPA(VTHWT,HAT,T2,31)+ RWA* SINTPA(VTHWT,HWT,T2,31))/
     1     RWAP1
      QADDP = 0.4536*NC*WC*(QADD/NC/WC+LMBDA1*THETA)/1055.0/RWAP1/NA/
     1     XS/PHI/WA
      HA = HAO + QADDP
      HH(3) = HAO+(1.0+RMRP*RMAP/RMAW)*QADDP-RMRP*RMAP/RMAW*(HA-HAP)+
     1     RMRP/RMAW*(1.0-RMAP)*HCOMB/2325.98
      H2 = HH(3)
       TT(3)=FUNCT3(H2)
           QAMAWR = (H2 - HH(2)) * 2325.98
           HH(4) = ZTP(ZH, TT(4), PP(4), NZDIM, NT, NP, VT, VP)
           SS(4) = ZTP(ZS, TT(4), PP(4), NZDIM, NT, NP, VT, VP)
           RHORHO(4) = ZTP(ZW, TT(4), PP(4), NZDIM, NT, NP, VT, VP) *
     1                 PP(4)/(R*TT(4))
      WRITE(IOUT,2500)ETAD,ETAG,ETAI
      WRITE(IOUT,2600)LMBDA1,LMBDA2
      WRITE(IOUT,2700)F,PAUXP1,DRP31,PCMAWR,QAMAWR,THETA,PMHD,PE
      WRITE(IOUT,3000)C,DK,DB
           RD(1) = 1.0
```

```
                    X(1) = 0.0
                    S(1) = SS(4)
                    SS(5) = SS(4)
              U(1) = U0
                    HH(5) = HH(4) - 0.5*U(1)*U(1)
                    H(1) = HH(5)
                    TT(5) = TZZ(S(1), ZS, H(1), ZH, NZDIM, NT, NP, VT, VP)
                    T(1) = TT(5)
                    P6T = PZT(S(1), ZS, T(1), NZDIM, NT, NP, VT, VP)
                    PP(5) = P6T
                    P(1) = P6T
                    SIGMA(1) = ZTP(ZSIGMA, T(1), P(1), NZDIM, NT, NP, VT, VP)
                    RHORHO(5) = ZTP(ZW, T(1), P(1), NZDIM, NT, NP, VT, VP) *
     1                     P(1)/(R*T(1))
                    RHO(1) = RHORHO(5)
                    MU(1) = ZTP(ZMU, T(1), P(1), NZDIM, NT, NP, VT, VP)
              K = K0
              VK(1) = K
              B(1) = B0
                    MUB(1) = B(1) * MU(1)
                    JSAFE = 5
                    P5S = PP(7) + 0.08
                    MG = PE / ((ZPZ(ZH, P5S, SS(4), ZS, NZDIM, NT, NP, VT, VP)
     1                 - HH(4)) * (0.05 - K))
              IF(MG .LE. 0.0)  MG = 1300.0 * PMHD / 1.3E9
                    KSAFE = JSAFE
        201         CONTINUE
                          DN = 0.0
                          IY = 1
                          IQ = 0
                          D4 = DSQRT(MG/(U(1)*RHO(1)))
                          RD4 = 1.0/D4
                          FXI = RHO(1)*U(1)*F*2.0/(B(1)**2*SIGMA(1)*D4)
                          N(1) = K*(1.0-K)/(1.0-K+FXI)
                          IV = 10
                          DP = 10.0
                          DP6T = 10.0
                          P6T = 10.0
                          JCNT = 1
C         ***                 REPLACE XP CALL , II = XP(P(I), 0) + 2
                    DO 290 I=1,NP
                    IF (P(1) .GE. VP(I))    GO TO 290
                               II = I
                               GO TO 295
        290           CONTINUE
                    II = NP
        295         CONTINUE
C                                   P6T LOOP
        301         CONTINUE
                          IF (MOD(IQ,2) .EQ. 0)  GO TO 305
                                P6T = P6T + DP6T
                                P(IY) = (P6T-PP(7))/DP6T*DP + P(IY-1)
                                DP = P(IY-1) - P(IY-2)
                                DN = N(IY-1) - N(IY-2)
                                FXI = FXIM - FXI
                                GO TO 315
        305               CONTINUE
```

```
                    IY = IY+1
                    II = II-1
C       ***   REPLACE PX CALL ,  P(IY) = PX(II, D)
                 P(IY) = VP(II)
    315          HS = ZPZ(ZH, P(IY), S(IY-1), ZS, NZDIM, NT, NP, VT,
      1                VP)
                 DHS = HS - H(IY-1)
                 TEMP = 2.D*C*DHS + U(IY-1)**2
                 IF (TEMP .GE. D.ODD)    GO TO 325
                    WRITE(IOUT, 2310)  TEMP, C
                    GO TO 500
    325          U(IY) = DSQRT(TEMP)
                 DPOLD = DP
                 DP = P(IY) - P(IY-1)
                 N(IY) = DN/DPOLD*DP + N(IY-1)
                 FXIM = 2.D*FXI
C                   VARY FXIM, CONVERGENCE BASED ON N
                 FACTX = 8.0*KP*(1.0+C)
                 FACTN = 0.5*RLH*(1.0+C)
                 FT2 = 2.0*F
C
    351       CONTINUE
                 H(IY) =  (FACTN*(N(IY)+N(IY-1))-C)*DHS + H(IY-1)
                 T(IY) = TPZ(P(IY), H(IY), ZH, NZDIM, NT, NP, VT, VP)
                 RHO(IY) = ZTP(ZW, T(IY), P(IY), NZDIM, NT, NP, VT, VP)
      1                *P(IY)/(R*T(IY))
                 SIGMA(IY) = ZTP(ZSIGMA, T(IY), P(IY), NZDIM, NT, NP,
      1                VT, VP)
                 RD(IY) = DSQRT(MG/(U(IY)*RHO(IY))) * RD4
                 K = VK(IY-1) + DK*DHS*RHO(IY)/101300.
                 VK(IY) = K
                 B(IY) = B(IY-1)
                 IF (P(IY) .LE. 2.0)  B(IY) = B(IY) +
      1             DB*DHS*RHO(IY)/101300.0
                 FXIOLD = FXI
                 FXI = FT2*RHO(IY)*U(IY) / (SIGMA(IY)*RD(IY)*D4*
      1                B(IY)**2)
                 FXIM = FXIM - FXIOLD + FXI
                 OLDNY = N(IY)
                 N(IY) = K*(1.0-K)/(1.0-K+FXI)
                 IF (DABS(OLDNY-N(IY)) .LT. 1.0E-5)    GO TO 355
                 JCNT = JCNT -1
                 IF (JCNT .LE. 1)  GO TO 355
              GO TO 351
    355          X(IY) = DP / ((K-0.5*FXIM-1.0) * (SIGMA(IY-1)+SIGMA(IY
      1             )) * (B(IY-1)**2 + B(IY)**2) * (U(IY-1)+U(IY))) *
      2             FACTX + X(IY-1)
                 S(IY) = ZTP(ZS, T(IY), P(IY), NZDIM, NT, NP, VT, VP)
                 MU(IY) = ZTP(ZMU, T(IY), P(IY), NZDIM, NT, NP, VT, VP)
                 MUB(IY) = MU(IY)*B(IY)
                 DN = N(IY) - N(IY-1)
                 JCNT = 4
                 IF (P(IY) .GT. PP(7))    GO TO 365
                    HH(7) = 0.5*U(IY)**2 + H(IY)
                    TSS = TZZ(S(IY), ZS, HH(7), ZH, NZDIM, NT, NP,
      1                   VT, VP)
                    PSS = PZT(S(IY), ZS, TSS, NZDIM, NT, NP, VT, VP)
```

```
                          P6TOLD = P6T
                          P6T = ETAD*(P5S-P(IY)) + P(IY)
                          DP6T = P6TOLD - P6T
  365               Q = IQ
                    IQ = 0
                    IF ((P6T-10.0*Q) .LE. PP(7))  IQ = 1
C                         CONVERGENCE TEST FOR P6T
                    IV = IV-IQ
                    IF (IV .LT. 0)  GO TO 375
                    IF (DABS(DP6T) .LT. 1.0D-10)  GO TO 375
                    IF (DABS(PP(7)-P6T) .LT. 0.005)  GO TO 375
              GO TO 301
  375 CONTINUE
                    WRITE(IOUT, 2300)  MG
                    OLDMG = MG
                    MG = PE*RLH/(HH(4)-HH(7))
                    IF (DABS(OLDMG/MG-1.0) .LT. 2.0D-4)  GO TO 405
                    KSAFE = KSAFE - 1
                    IF (KSAFE .LE. 0)  GO TO 405
              GO TO 201
C                         END OF KSAFE (FXI) LOOP
  405 CONTINUE
              HH(6) = H(IY)
              SS(6) = S(IY)
              PP(6) = P(IY)
              TT(6) = T(IY)
              RHORHO(6) =  RHO(IY)
              H5SP = ZPZ(ZH, P5S, SS(4), ZS, NZDIM, NT, NP, VT, VP)
              ETAMHD = PE / (MG*(HH(4)-H5SP))
              TT(7) = TPZ(PP(7), HH(7), ZH, NZDIM, NT, NP, VT, VP)
              SS(7) = ZTP(ZS, TT(7), PP(7), NZDIM, NT, NP, VT, VP)
              RHORHO(7) = ZTP(ZW, TT(7), PP(7), NZDIM, NT, NP, VT, VP) *
     1                         PP(7)/(R*TT(7))
              TT(8) = TT(7)
              SS(8) = SS(7)
              RHORHO(8) = RHORHO(7)
              HH(8) = HH(7)
              IF (RMAP .LE. 1.0D-30)  GO TO 425
  411               DHA =(SINTPA(VTHWT, HAT, TT(8), 31) +
     1                    SINTPA(  VTHWT,HWT,TT(8),31)*RWA)/RWAP1-HAMB
                    HH(8) = (RMAP*DHA + HH(7)) / (RMAP+1.0)
                    TT8NEW = SINTPA(V8H, V8T, HH(8), N8T)
                    IF (DABS(TT(8)-TT8NEW) .LE. 0.1)  GO TO 415
                          TT(8) = TT8NEW
              GO TO 411
  415               TT(8) = TT8NEW
                    SS(8) = SINTPA(V8T, V8S, TT(8), N8T)
                    RHORHO(8) = SINTPA(V8T, V8W, TT(8), N8T) * PP(8)/
     1                         (R*TT(8))
  425         HH(9) = HH(8) - QAMAWR*RMAW/RMGP
              TT(9) = SINTPA(V8H, V8T, HH(9), N8T)
              SS(9) = SINTPA(V8T, V8S, TT(9), N8T)
              RHORHO(9) = SINTPA(V8T, V8W, TT(9), N8T) * PP(9)/(R*TT(9))
              DQI = PE - PMHD
              Q50 = MG * RMGP * (HH(9)-HH(10))
              QC = RMC * THETA * MG
              Q51 = LMBDA1 * QC
```

```
                  QS2 = LMBDA2 * PE
                  PC = PCMAWR * RMAW * MG
                  ETAS = 0.39
                  DO 480  KY=1,7
                       ETAS = ETAS + 0.01
                       ETASF = 0.995 * ETAS
                       RETAG = 1.0/ETAG
                       VETAS(KY) = ETAS
                       PS(KY) = ((1.0-ETAG)*PC -QS0 - QS1 - QS2 - DQI)*ETASF/
         1                  ((RETAG-1.0)*ETASF-1.0)
                       PSE(KY) = PS(KY) - PC*ETAG
                       DQG(KY) = PSE(KY) * RETAG - PSE(KY)
                       QS(KY) = 0.995 * (DQG(KY)+QS0+QS1+QS2+DQI)
                       PO(KY) = (1.0-PAUXP1)*PSE(KY) + PMHD
                       ETA(KY) = PO(KY)/QC
    480           CONTINUE
C                      OUTPUT
                  LPP3 = 10.0*PP(4)
                  LK = 100.0*K
                  YY = 10.0*DRP31
                  IV = YY + 0.5
                  KY = 10.0*(YY-IV)
                  LU0 = U(1)
                  LC = 100.0*C + 0.5
                  LTT3 = TT(4)
                  LB0 = B(1)
                  RMRPMG = RMRP*MG
                  RMOXMG = RMOX*MG
                  RMAWMG =  RMAW *MG
             RMAMG = RMA*MG
                  RMGPMG=  RMGP *MG
                  RMAPMG = RMAP *MG
                  RMCMG  = RMC *MG
                  RMCWMG = RMCW *MG
                  RMSMG  = RMS* MG
                  RMWCMG = RMWC *MG
                  RMWSMG = RMWS *MG
                  RMARP = RMAWMG+RMRPMG
                  RGMAMG = MG +RMAPMG
             WRITE(IOUT,2800) PP(1), TT(1),PP(2),TT(2),PP(3),TT(3)
             WRITE(IOUT,2900) PP(4),TT(4), PP(7),TT(7),PP(8),TT(8),
         1        PP(9),TT(9), PP(10),TT(10)
                  WRITE(IOUT, 2420)
             WRITE(IOUT,2424)PAMB,TAMB,HAMB,RMAWMG
             WRITE(IOUT,2423) LTAB1(1),PP(1),TT(1),HH(1),RMAWMG
             WRITE(IOUT,2423) LTAB1(2),PP(2),TT(2),HH(2),RMARP
             WRITE(IOUT,2423) LTAB1(3),PP(3),TT(3),HH(3),RMARP
             WRITE(IOUT,2425)(LTAB1(J),PP(J),TT(J),HH(J),RHORHO(J),SS(J),MG,
         1        J=4,6 )
             WRITE(IOUT,2425)LTAB1(7),PP(7),TT(7),HH(7),RHORHO(7),SS(7),MG
             WRITE(IOUT,2425)(LTAB1(J),PP(J),TT(J),HH(J),RHORHO(J),SS(J),
         1        RGMAMG,J=8,10)
                  WRITE(IOUT, 2430)
                  JY = IY - 1
                  IF (JY .LE. 0)    GO TO 485
                  WRITE(IOUT,2435)(LTAB2(J),X(J), RD(J), P(J), T(J), SIGMA(J),
         1        H(J), RHO(J), U(J), MU(J), S(J), N(J), MUB(J),
```

```
     2      VK(J), B(J), J=1,JY)
  485       J = IY
            WRITE(IOUT, 2435)  LAB5, X(J),RD(J), P(J), T(J), SIGMA(J),
     1            H(J), RHO(J), U(J), MU(J), S(J), N(J), MUB(J)
     2      , VK(J), B(J)
            V = 0.5 * D4 * (RD(IY)+1.0)
            XVD4 = (X(IY) + V) * V
            D5 = RD(IY) * D4
            A4 = D4 * D4
            A5 = D5*D5
            WRITE(IOUT, 2440)  MG, RMAMG, RMAWMG, RMGPMG, RMAPMG,
     1            RMCMG, RMCWMG, RMSMG, RMWCMG, RMWSMG, XVD4, D4, D5,
     2            A4, A5, WC, ETAMHD
            DO 490 KY=1,7
                  PSETAG(KY) = PSE(KY)/ETAG
                  PSEMHD(KY) = PSE(KY) + PMHD
                  PSEPAU(KY) = PSE(KY) * PAUXP1
                  CETA(KY) = 3412.16/ETA(KY)
  490       CONTINUE
      WRITE(IOUT,2460)
      WRITE(IOUT,2480)(VETAS(J),J=1,7)
      WRITE(IOUT,2470)(QS0,J=1,7),(QS1,J=1,7)
     1            , (QS2,J=1,7), (DQI,J=1,7), (DQG(J),J=1,7),
     2            (QS(J),J=1,7), (PS(J),J=1,7),
     3            (PSETAG(J),J=1,7), (PC,J=1,7), (PSE(J),J=1,7),
     4            (PMHD,J=1,7), (PSEMHD(J), J=1,7), (PSEPAU(J),J=1,7),
     5      (P0(J),J=1,7),(CETA(J),J=1,7)
      WRITE(IOUT,2490)(ETA(J),J=1,7)
  499 CONTINUE
  500 CONTINUE
      GO TO 101
 2100 FORMAT(8F10.0)
 2110 FORMAT(I2)
 2120 FORMAT(6F10.0)
 2200 FORMAT(1H1, 12A6 / 1H , 12A6)
 2210 FORMAT(1H0, 6HPHI = , F6.2, 5X, 6HPMHD =, 1PE12.5, 5X, 7HTCOMB =,
     1      0PF8.1, 5X, 7HPCOMB =, F8.3, 5X, 4HU0 = , F8.2,
     2      /,1X,7HRMKP = ,F10.3,5X,7HRMOX = ,F10.3)
 2220 FORMAT(1H0, 6HRMAW =, 1PE14.6, 5X, 6H RMC =,E14.6, 5X, 6H RMS =,
     1      E14.6 /1H , 6HRMWC =, E14.6, 5X, 6HRMWS =, E14.6, 5X,
     2      6HRMCW =, E14.6, 5X, 6HRMAP =, E14.6)
 2230 FORMAT(1H0, 4HWC = , F8.2, 5X, 4HNC =, F8.4, 5X, 4HNA = F8.4,
     1      5X, 8HHHVDAF =, 1PE14.6 /1H ,  4HXS =, E14.6, 5X,
     2      10HT(STACK) =, E14.6)
 2240 FORMAT(1H0, 24H**ITERATION ON T1 FAILED)
 2250 FORMAT(1H0, 24H**ITERATION ON T2 FAILED)
 2300 FORMAT(1H , 6H    MG, 1PE13.6)
 2310 FORMAT(1H0/1H0, 42H** CALCULATION OF FLUID VELOCITY INVOLVES ,
     1      22HTAKING SQUARE ROOT OF , 1PE12.5, 1H. /1H , 5X,
     2      21HCURRENT VALUE OF C = , E12.5,
     3      43H. HIGHER VALUES OF C WILL ALSO BE IGNORED. )
 2400 FORMAT(1H , 2X, 2(1H-, I2), 1H-, 2I1, 1H-, I3, 1H-, I2, 1H-, I4,
     1      1H-, I1)
 2410 FORMAT(1H0, 3X, 5HINPUT/1H0, 10X, 5H PHI , F5.2, 7H CYCLE , I2,
     1      4H P3 , F6.2, 4H T3 , F5.0,12H (P1-P3)/P3 , F6.2,
     2      3H U , F4.0, 3H C , F5.2, 9H LAMBDA1 , F5.2, 9H LAMDA2  ,
     3      F5.2 /1H0, 3H K , F5.2, 4H PE , 1PE8.1, 4H P6 , 0PF6.2,
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     4        6H ETAU , F5.2, 3H F , F6.3, 4H T8 , F4.0, 3H B , F2.0,
     5        8H PC/MAW , F7.0, 4H P1 , F6.2, 4H T1 , F5.0, 4H P2 , F6.2
     6        /1H0, 4H T2 , F5.0, 6H ETAG , F6.3, 6H ETAI , F6.3,
     7        10H PAUX/PSE , F6.3, 8H QA/MAW , F7.0, 7H THETA , 1PE10.3
     8        / (1H0, 6H ETAS , 0PF6.2, 6(1H, F6.2)))
 2420 FORMAT(1H0,//,11X,       10HGAS STATES  ,//,2X,6HPOINT,5X,1HP,
     1        10X, 1HT, 10X, 1HH, 10X, 3HRHO, 9X, 1HS,9X,1HM,//)
 2425 FORMAT(1H , 2X, A3, 4X, F6.3, 4X, F5.0, 4X, F9.0, 4X, F7.4,
     1        4X, F7.1, 4X, F8.4)
 2430 FORMAT(1H1,12X, 16HGENERATOR DESIGN  ,1H0,7H  POINT,
     1        2X, 6HLENGTH, 4X, 4HD/D4, 5X, 1HP, 7X, 1HT, 4X, 5HSIGMA,
     2        7X, 1HH, 7X, 3HRHO, 6X, 1HU, 7X, 2HMU, 8X, 1HS, 7X, 1HN,
     3        6X, 4HMU*B, 5X, 1HK, 5X, 1HB)
 2435 FORMAT(1H , 3X, A3, 2X, F7.3, 2X, F6.3, 2X, F6.3, 2X, F5.0,
     1        2X, F6.3, 2X, F8.0, 2X, F7.4, 2X, F6.1, 2X, F7.4, 2X, F7.1,
     2        2X, F6.4, 2X, F7.4, F6.3, F6.2)
 2440 FORMAT(1H0, 3X, 25HMASSFLOWS, GENERATOR AREA
     1        /1H , 5X, 4H MG , F8.3, 4H MA , F8.3, 4H MAW, F8.3,
     2             4H MG*, F8.3, 4H MA*, F8.3
     3        /1H , 5X, 4H MC , F8.3, 4H MCW, F8.3, 4H MS , F8.3,
     4             4H MWC, F8.3, 4H MWS, F8.3
     5        /1H , 6X, 15H (X + DM) = DM , F9.5, 3H D4, F7.4,
     6             3H D5, F7.4, 3H A4, F7.4, 3H A5, F7.4
     7        /1H , 6X, 12H THETA * MC , 1PE12.5, 8H ETAMHD , 0PF8.5)
 2423 FORMAT(1H ,2X,A3,4X,F6.3,4X,F5.0,4X,F9.0,22X,4X,F8.3)
 2424 FORMAT(1H ,2X,3H0  ,4X,F6.3,4X,F5.0,4X,F9.0,22X,4X,F8.3)
 2460 FORMAT(1H0,13X,          19HOVERALL PERFORMANCE )
 2470 FORMAT(1HH,4X,8HQS0           ,3X,7E14.5
     3        /1H , 4X, 8HQS1         ,3X,7E14.5
     4        /1H , 4X, 8HQS2         ,3X,7E14.5
     5        /1H , 4X, 8HDQ1         ,3X,7E14.5
     6        /1H , 4X, 8HDQG         ,3X,7E14.5
     7        /1H , 4X, 8HWS          ,3X,7E14.5
     8        /1H , 4X, 8HPS          ,3X,7E14.5
     9        /1H , 4X, 8HPT          ,3X,7E14.5
     1        /1H , 4X, 8HPC          ,3X,7E14.5
     2        /1H , 4X, 8HPSE         ,3X,7E14.5
     3        /1H , 4X, 8HPMHD        ,3X,7E14.5
     4        /1H , 4X, 8HPSE+PMHD,3X,7E14.5
     5        /1H , 4X, 8HPAUX        ,3X,7E14.5
     6        /1H , 4X, 8HP           ,3X,7E14.5
     7        /1H , 4X, 8HHR          ,3X,7E14.5)
 2480 FORMAT(1H0     ,11HSTEAM PLANT   ,/ ,1X,10HEFFICIENCY ,7(2X,F12.2))
 2490 FORMAT(1H0     ,7HOVERALL ,/,1X,11HEFFICIENCY   ,2X,0PF12.4,
     1        6(2X,F12.4))
 2500 FORMAT(1H0,1X,12HEFFICIENCIES  /,5X,8HDIFFUSER ,12X,F10.4 /
     1        ,5X,18HROTATING GENERATOR ,2X,F10.4 ,/
     2        ,5X,14HDC/AC INVERTER ,6X,F10.4  )
 2600 FORMAT(1H0,1X,19HHEAT TRANSFER RATIO , / , 6X,
     3        37HFROM COMBUSTER TO SUBPOSED PLANT    = , F8.3, / ,6X,
     4        37HMHD GENERATOR TO SUBPOSED PLANT     = , F8.3 )
 2700 FORMAT(//,6X,29HFRICTION FACTOR IN MHD DUCT = ,1X,F8.4  ,/,6X,
     5        10HPAUX/PSE =,1X,F8.4,/,6X,18H(P1-PCOMB)/PCOMB =,1X,F8.4,/,
     6        6X, 8HPC/MAW =,1X,1PE12.5, /,6X, 8HQA/MAW =,1X,E12.5, /,6X,
     7        7HTHETA =,4X,E12.5 ,/, 6X,6HPMHD =,4X,E12.5 , /, 6X,
     8        4HPE =,5X ,E12.5 )
 2800 FORMAT(1H0,1X,8HAIR SIDE ,8X,5HTOTAL ,3X,8HPRESSURE  ,3X,
```

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          $       11HTEMPERATURE   ,
          9        //,4X,16HCOMPRESSOR INLET , 3X,F10.4, 4X,F10.4 , / ,
          A       4X,17HCOMPRESSOR OUTLET ,2X,F10.4, 4X,F10.4 , /,
          B       4X,18HAIR PREHEATER EXIT ,1X,F10.4, 4X,F10.4 )
     2900 FORMAT(1H0,1X,8HGAS SIDE  ,14X,5HTOTAL  ,3X,8HPRESSURE  ,3X,
          %       11HTEMPERATURE   ,
          C        //, 4X,14HMHD DUCT INLET,12X,F10.4,4X,F10.4 , / , 4X,
          D       13HDIFFUSER EXIT ,13X,F10.4 ,4X,F10.4 , /, 4X,
          E       13HINJECTOR EXIT ,13X,F10.4 ,4X,F10.4 , /, 4X,
          F       18HAIR PREHEATER EXIT , 8X, F10.4 , 4X, F10.4 , /, 4X,
          G       25HBOTTOMING HEAT EXCH. EXIT ,1X ,F10.4 ,4X,F10.4 )
     3000 FORMAT(1H0,4X,3HC = ,1X,F8.4,/,5X,4HDK = ,1X,F8.4,/,5X,4HDB = ,
          1       1X,F8.4 )
     9999 CALL EXIT
          END
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@PRT,SC PCF.FUNCT2

6. PRELIM

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*MHDDUCT.PRELIM
      SUBROUTINE PRELIM(RMAW,RMC,RMS,RMWC,RMWS,RMCW,RMAP,RMRP,RMOX,
     1 HHVDAF,XS,FMW,A1,DAM,ITERAT)
C     HF ARE HEATS OF FORMATION OF SEED COMPOUNDS
      IMPLICIT DOUBLE PRECISION(A-H,O-Z)
      DIMENSION AMW(4),SMW(4),HF(4)
      DIMENSION HA(24)
C     AMW ARE THE MW OF K2.CS2 AND SMW ARE MOL. WT. OF SEED COMPOUNDS
      REAL MAMC, MAMMC,MWCMC,MWSMC, MSMC, MOXMC,MGMC,MGMCP
      DATA AMW/78.200,78.200,265.8200,265.8200/
     1 SMW/138.200,174.2600,325.8300,361.8900/
     2 HF/-274960.D0,-338620.D0,-267400.D0,-339380.D0/
   10 FORMAT ( )
    8 FORMAT(    12A6/12A6)
    9 FORMAT(1H1)
      WRITE(6,9)
      IF(ITERAT.GE.1)GO TO 20
      READ(5, 8           )(HA(I),I=1,24)
C     READ IN FUEL COMPOS. AS WT. PERCENT OF CARBON C , HYDROGEN H,
C     OXYGEN FO2, NITROGEN FN2, SULFUR S,MOISTURE,ASH
C     HHV AND SENSIBLE HEAT OF FUEL ON AS RECEIVED BASIS
      READ(5,10) C, H, FO2, FN2, S,     FH2O, ASH, HHV, FSH
C     READ PERCENT MOIST AFTER DRYING - AS FIRED
      READ(5,10)DFH2O
C     READ IN SEED TYPE 1= K2CO3, 2 =K2SO4, 3= CS2CO3 4= CS2SO4 AND WT.,
C     MOLES OF WATER TO SEED MOLES     IN SEED SOLUTION AND WT. PERCENT
C     OF SEED IN PRODUCTS
      READ(5,10) LASC, RWS, R
      R=R/100.
C     READ IN WT. PER CENT OF O2 IN AIR + O2 AND MOISTURE IN AIR SUPPLY
      READ(5,10) AO2, RWA
C     CALCULATE COMP. SUBSCRIPTS OF CHEM. FORM. DRY ASH FREE FUEL
      CI = C / 12.011
      HI = H/1.008 / CI
      FOI   = FO2 / 16. / CI
      FNI = FN2 / 14.008 / CI
      SI = S/32.066 / CI
      CI = 1.
C     FMW IS MW OF DAF COAL
      FMW = 12.01 + 1.008*HI + 16.* FOI + 14.008* FNI + 32.066* SI
      HHVDAF =HHV*100./(100.-FH2O-ASH)
C     CALCULATE HEATING VALUE IN KCAL/KG. MOLE
      HVPM = HHVDAF*2325.43/4184.*FMW
      HOF = HVPM - CI* 94040. - HI *0.5 * 68317. - SI * 69300.
C     CALCULATE H2O ENTERING AS MOISTURE IN THE COAL AS MOLE RATIO TO
C     DRY ASH FREE FUEL LAMBDA
      RLA =DFH2O /(100. -DFH2O - ASH) * FMW /18.016
C     CALCULATE MOLE FRACTION OF SEED COMPOUND IN FUEL MIX. A3
      C = 1./(1. + RLA)
      ALF =RWS * SMW(LASC) / 18.016
      D =(1. + ALF) * C
      A3 = R*C* (FMW + RLA * 18.016 )/(AMW(LASC) +R*(D* FMW +D*RLA*
     1     18.016 - SMW(LASC) - ALF *18.016 ))
C     CALCULATE MOLE FRACTION OF COMBUSTIBLE A1 AND WATER A2 IN FUEL
      A1 = C - D* A3
      A2 = A1* RLA + ALF*A3
   11 FORMAT(///12A6/12A6)
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      WRITE(6,11)(HA(I),I=1,24)
   13 FORMAT(28X,16HFUEL COMPOSITION/2X,11HCOMB. MOLES,4X,6HCARBON,5X,
     1       8HHYDROGEN,7X,6HOXYGEN,7X,8HNITROGEN,3X,6HSULFUR )
      WRITE(6,13)
C     PRINT OUT THREE LINES OF SECOND INPUT SHEET FOR 2502
      HFW = -68317.
      WRITE(6,10) A1,CI,HI,FOI,FNI,SI.
   14 FORMAT(3X,25HCOMB. HEAT OF FORMATION= ,F9.1,3X,14HCOMB. DAF HHV=,
     1 F9.1/3X,19HMOIST. MOLE FRACT.=,F9.6,3X,23HSEED COMPOUND MOLE FRAC
     2       ,3HT.=,F9.6/3X,17HCOMB. MOL WT(WC)=,F9.4)
      WRITE(6,14) HOF,HHVDAF,A2,A3,FMW
C     CALC. MOLES OF MOIST. BROUGHT IN WITH EACH MOLE OF DRY AIR EPS
      EPS = RHA /18.016 * 28.9703/ 100.
      A = 1. /(1.+EPS)
      B = (1. + ALF) * A
C     CALCULATE MOLS OF O2 PER MOL OF DRY AIR ROA
      AO2=AO2/100.
      ROA=(AO2-.2319)/(1.-AO2)*28.9703/32.
      E = (1. + EPS) /(1. + ROA + EPS )
C     CALC. MOLES OF SEED COMPOUND FOR EACH MOL OXIDANT B7
      B7 = R*E*A*(28.9703 + EPS*18.016 +ROA*32.)/(AMW(LASC) + R*(E*B*
     1       (28.9703    +EPS*18.016 + ROA * 32.) -SMW(LASC) -ALF*18.016))
C     CALC. DRY AIR MOLES PER OXIDANT MOLE - DAM
      DAM = E * (A - B* B7)
C     CALC. INPUTS FOR OXIDANT IN 2502
      B1 = 0.2099* DAM
      B2 = 0.7803 * DAM
      B3 = 0.0003 * DAM
      B4 = 0.0095 * DAM
      B8 =ROA * DAM
C     THE MOIST. ASSOC. WITH AIR IS SEPARATED FROM THAT WITH SEED SINCE
C     THEY ENTER IN DIFFERENT PHASES
      B5 = EPS * DAM
      B6 = ALF * B7
C     PRINT OUT NOS. NEEDED FOR 2502
      HCO2 = -94040.
      HB5 = -57760.
   15 FORMAT(23X,19HOXIDANT COMPOSITION /12H O2 MOL FRAC, 2X,11HN2 MOL F
     1RAC,2X,12HCO2 MOL FRAC, 2X,11HA MOL FRAC, 2X,14HSEED MOL FRAC  )
      WRITE(6,15)
      WRITE(6,10) B1,B2,B3,B4,B7
   16 FORMAT(3X,22HMOL FRACT OF VAP H2O= ,F9.6,5X,22HMOL FRACT OF LIQ H2
     1O= ,F9.6/3X,23HMOL FRAC OF O2 ENRICH.= ,F9.6)
      WRITE(6,16) B5,B6,B8
C     BEGIN CALC. OF MASS FLOW RATE RATIOS
C     XS IS THE STOICHIOMETRIC MOLES OF OXIDANT PER MOLE OF FUEL
C     FIRST XS GIVES O2 REQUIRED BY COMB. + CARBONATE SEED IN FUEL+ OXID.
      SM = 0.
      IF (MOD(LASC,2).EQ.0.) SM = 0.5
      XS=(A1*(CI + 0.25*HI - 0.5*FOI + SI) - SM*A3)/(B1 + SM*B7)
C     CALC. MOLECULAR WT OF OXIDANT  OMW
            OMW = DAM   *28.9703 + (B5 +B6)* 18.016 + B7*SMW(LASC)+B8*32
   17 FORMAT(3X,22HMOL FRAC OF  AIR (NA)=,F9.6,3X,19HMOL FRAC OF O2 (NOX
     1       ,2H)=,F9.4)
      WRITE(6,17) DAM,B8
C     RMRP IS THE  MASS RATIO OF RECIRC. PRODS. TO DUCT FLOW
      READ(5,10)  PHI, RMRP
```

```
      C     CALCULATE MASS FLOW RATIOS FOR INPUT EQUIV. RATIOS
      C     RMAP IS RATIO OF SUPPLEMENTARY AIR TO MG
         20 RMAP =0.
      C     CALCULATE MOLES OF SEED SUPPLIED FOR EACH MOLE OF DAF COMBUSTIBLE
            SMCM =(A3 + XS*PHI* H7) /A1
            MWSMC = SMCM *R.5 *18.016 /FMW
            MSMC = SMCM *SMW(LASC)/FMW
            MAMC= DAM*28.9703  /FMW * XS* PHI /A1
            MWCMC = RLA*18.016/FMW
      C      MOXMC MAY NOT BE CORRECT RMRP*RMAP>0.  2/20/75  DQH
            MOXMC=ROA*32./28.9703 * MAMC
            MAWMC =(1.+ RWA/100.)* MAMC
            MGMCP=(1.+ MAWMC + MWCMC + MWSMC + MSMC + MOXMC)/(1.-RMRP)
            IF (PHI .LT. 1.) RMAP=(1.05-PHI)/PHI* MAWMC/MGMCP/(1.-RMRP)
            DO 40 K=1,15
            RMAW=MAWMC/MGMCP
            MAWMC =(1.+ RWA/100.)* MAMC    - RMRP*RMAP*MGMCP
            MGMC =(1.+ MAWMC + MWCMC + MWSMC + MSMC + MOXMC)/(1.-RMRP)
            IF (ABS((MGMCP-MGMC)/MGMC).LE..0001) GO TO 42
         40 MGMCP=MGMC
         41 FORMAT(1H0,'MGMC ITERATION HAS NOT CLOSED')
            WRITE(6,41)
         42 RMC=1./MGMC
            RMAW=RMC*MAWMC
            RMS =RMC*MSMC
            RMWC=RMC*MWCMC
            RMWS=RMC*MWSMC
            RMCW=RMC+RMWC
            RMOX=RMC*MOXMC
            IF(ITERAT.LE.0)GO TO 39
            IF(ITERAT.LT.10)GO TO 200
         39 CONTINUE
            WRITE(6,10)MGMC,MAWMC
         22 FORMAT(25X,21HMASS FLOW RATE RATIOS/3X,4HPHI=,F7.3,5X,9HMASS=,F9.5)
            WRITE(6,22) PHI,XS
            WRITE(6,23)
         23 FORMAT(5X,4HRMOX,9X,4HRMRP,9X,4HRMAP)
            WRITE(6,10)RMOX,RMRP,RMAP
         24 FORMAT(5X,4HRMAW,9X,3HRMC,10X,3HRMS,9X,4HRMWC,10X,4HRMWS,9X,4HRMCW
           1       )
            WRITE(6,24)
        100 WRITE(6,10) RMAW, RMC, RMS, RMWC, RMWS, RMCW
        200 CONTINUE
            RETURN
            END
```

RT,SC   PCF.PWRE

7. SETUP

```
EI.MHDDUCT.SETUP
      SUBROUTINE SETUP(PHI)
      IMPLICIT DOUBLE PRECISION (A-H, O-Z)
      INTEGER HEAD
      COMMON        NT, NP, HEAD(24),
     1      VT(50), VP(50), ZMU(50,50), ZW(50,50), ZH(50,50),
     2      ZSIGMA(50,50), ZS(50,50), ZQADD(50,50), SC(10,10), SB(10,10)
      READ(4)  HEAD
      READ(4) NT, NP, NFUEL, NAGENT, PHI
      READ(4)  (VT(I), I=1,NT)
      READ(4)  (VP(I), I=1,NP)
      READ(4) ((SC(I,J),I=1,10), J=1,NFUEL), ((SB(I,J),I=1,10),J=1,
     1            NAGENT)
      READ(4)  ((ZMU(I,J),I=1,NT), J=1,NP)
      READ(4)  ((ZQADD(I,J),I=1,NT), J=1,NP)
      READ(4)  ((ZW (I,J),I=1,NT), J=1,NP)
      READ(4)  ((ZH (I,J),I=1,NT), J=1,NP)
      READ(4)  ((ZSIGMA(I,J),I=1,NT), J=1,NP)
      READ(4)  ((ZS (I,J),I=1,NT), J=1,NP)
      RETURN
      END
```

PRT,SC PCF.SINTPA

8. BLOCKDATA

```
@I.MHDDUCT.BLOCKDATA
      BLOCK DATA
      IMPLICIT DOUBLE PRECISION (A-H, O-Z)
      COMMON /CONT2/ HAT(31), HWT(31),VTHWT(31),RWA,RWAPI,H2,HOXT(31),
     1       VBH(50),RMOX,RMAW,RMRP,VBT(50),NBT
      DATA  HAT /        -2.19580D1,  1.99200D0,  2.59820D1,  5.00720D1,
     1       7.43220D1,  9.87720D1,  1.23492D2,  1.48502D2,  1.73812D2,
     2       1.99452D2,  2.25412D2,  2.51682D2,  2.78252D2,  3.05102D2,
     3       3.32222D2,  3.59602D2,  3.87222D2,  4.15062D2,  4.43102D2,
     4       4.71332D2,  4.99732D2,  5.28292D2,  5.57002D2,  5.85862D2,
     5       6.14842D2,  6.43962D2,  6.73192D2,  7.02542D2,  7.31992D2,
     6       7.61532D2,  7.91172D2 /
      DATA  HWT /        -4.06193996451D1,  3.68566252213D0,
     1       4.82405026637D1,    9.32393925390D1,   1.38837749555D2,
     2       1.85129934279D2,    2.32199206038D2,   2.80106621669D2,
     3       3.28863282414D2,    3.78491390764D2,   4.29018699623D2,
     4       4.80450760212D2,    5.32820875665D2,   5.86134596800D2,
     5       6.40397474240D2,    6.95598406750D2,   7.51742944940D2,
     6       8.08797765080D2,    8.66740724690D2,   9.25566213140D2,
     7       9.85240946720D2,    1.04572607105D3,   1.10701048490D3,
     8       1.16904978330D3,    1.23181621314D3,   1.29527092007D3,
     9       1.35938615098D3,    1.42413915275D3,   1.48948717230D3,
     .       1.55542855773D3,    1.62191390409D3 /
C  ENTHALPIES ARE ADJUSTED TO ZERO AT 32F
      DATA  HOXT/        -2.01889D1, 1.59544D0, 2.35361D1, 4.57548D1,
     1       6.83392D1, 9.13361D1, 1.14761D2, 1.38605D2, 1.62836D2,
     2       1.87433D2, 2.12355D2, 2.37564D2, 2.63046D2, 2.88768D2,
     3       3.14711D2, 3.40846D2, 3.67155D2, 3.93621D2, 4.20236D2,
     4       4.46993D2, 4.73880D2, 5.00886D2, 5.28011D2, 5.55246D2,
     5       5.82586D2, 6.10033D2, 6.37583D2, 6.65233D2, 6.92983D2,
     6       7.20830D2, 7.48777D2 /
      END
```

PRT,SC PCF.CHARABS

9. BISECT

```
EI•MHDDUCT•BISECT
      FUNCTION BISECT(F, X, AA, BB, EPS, ERROR)
      IMPLICIT DOUBLE PRECISION (A-H, O-Z)
      LOGICAL ERROR
C             BISECTION METHOD TO FIND F(X) = 0
      A = AA
      FA = F(A)
      B = BB
      FB = F(B)
      ERROR = .FALSE.
   15 IF(SIGN(1.0,FA)*FB .LE. 0.0)  GO TO 25
            ERROR = .TRUE.
            RETURN
   25 DX = 0.5*(B-A)
      C = A + DX
      X = C
      FC = F(C)
      BISECT = FC
      IF (ABS(DX) .LE. EPS  .OR.  FC .EQ. 0.0)  GO TO 45
            IF (SIGN(1.0,FC)*FA .GE. 0.0)  GO TO 35
                  B = C
                  FB = FC
                  GO TO 15
   35       CONTINUE
                  A = C
                  FA = FC
                  GO TO 15
   45 T = ABS(FC)
      IF (T .LE. ABS(FA)) GO TO 55
            X = A
            BISECT = FA
            RETURN
   55 IF (T .LE. ABS(FB))  RETURN
      X = B
      BISECT = FB
      RETURN
      END
```

PRT,SC PCF•BLOCKDATA

10. FUNCT2

```
.I•MHDDUCT•FUNCT2
      DOUBLE PRECISION FUNCTION FUNCT2(T2)
      IMPLICIT DOUBLE PRECISION (A-H, O-Z)
      COMMON /COMT2/ HAT(31), HWT(31),VTHWT(31),RWA,RWAP1,H2,HOXT(31),
     1         V8H(50),RMOX,RMAW,RMRP,V8T(50),N8T
      COMMON /REPR/HSTACK,TSTACK,PP(10),HRE
      H = (SINTPA(VTHWT,HAT,T2,31)+RWA*SINTPA(VTHWT,HWT,T2,31))/RWAP1
     1     +SINTPA(VTHWT,HOXT,T2,31)*RMOX/RMAW
     2     +SINTPA(   V8T,   V8H,T2,N8T)*RMRP/RMAW/2325.98
      FUNCT2 = H - H2
      RETURN
      END

?RT,SC  PCF.FUNCT3
```

11. FUNCT3

```
EI.MHDDUCT.FUNCT3
                DOUBLE PRECISION FUNCTION FUNCT3(H2)
                IMPLICIT DOUBLE PRECISION (A-H, O-Z)
                LOGICAL ERR
                T2 = 200.
                DHREF = FUNCT2(T2)
                EPS = 1.0E-3
                K=300
                DO 100 I=K,3000,100
                T2 = I
                DH = FUNCT2(T2)
                IF (DHREF*DH .LE. 0.0)   GO TO 115
            100 CONTINUE
            105 PRINT 810
                CALL EXIT
            115 CONTINUE
                TT = T2
                DH = BISECT(FUNCT2, T2, TT-100., TT, EPS, ERR)
                IF (ERR)  GO TO 105
                H = H2 + DH
                FUNCT3=T2
            810 FORMAT(1H0, 'NO SOLUTION')
                RETURN
                END

PRT.SC  PCF.MAIN
```

12. SINTPA

```
EI.MHDDUCT.SINTPA
          DOUBLE PRECISION
         1FUNCTION SINTPA (XT, YT, X, N)
          IMPLICIT DOUBLE PRECISION (A-H, O-Z)
          DIMENSION XT(1), YT(1)
C   SINGLE VARIABLE INTERPOLATION (LAGRANGE 3 POINT METHOD)
C                    X IN ASCENDING ORDER.
          NM1 = N-1
          K = 2
          IF (XT(2).GT.X) GO TO 40
          K = NM1
          IF (XT(K).LE.X) GO TO 40
          L = 2
       10 I = K-L
          IF (I.LE.1) GO TO 40
          J = (K+L)/2
          IF (XT(J)-X) 20,30,30
       20 L = J
          GO TO 10
       30 K = J
          GO TO 10
       40 CONTINUE
          Y1 = YT(K-1)
          Y2 = YT(K)
          Y3 = YT(K+1)
          X1 = XT(K-1)
          X2 = XT(K)
          X3 = XT(K+1)
          Z1 = X-X1
          Z2 = X2-X
          Z3 = X3-X1
          SINTPA = Y1 + (1.0+Z2/Z3)*Z1*(Y2-Y1)/(X2-X1) - (Y3-Y2)/(X3-X2)*
         1     Z2*Z1/Z3
          RETURN
          END

PRT,SC PCF.SINTPD
```

13. SINTPD

```
EI*MHDUUCT.SINTPD
          DOUBLE PRECISION
         1FUNCTION SINTPD (XT, YT, X, N)
          IMPLICIT DOUBLE PRECISION (A-H, O-Z)
          DIMENSION XT(1), YT(1)
      C  SINGLE VARIABLE INTERPOLATION (LAGRANGE 3 POINT METHOD)
      C                        X IN DESCENDING ORDER
          NM1 = N-1
          K = 2
          IF (XT(2).LT.X) GO TO 40
          K = NM1
          IF (XT(K).GE.X) GO TO 40
          L = 2
       10 I = K-L
          IF (I.LE.1) GO TO 40
          J = (K+L)/2
          IF (XT(J)-X) 30,30,20
       20 L = J
          GO TO 10
       30 K = J
          GO TO 10
       40 CONTINUE
          Y1 = YT(K-1)
          Y2 = YT(K)
          Y3 = YT(K+1)
          X1 = XT(K-1)
          X2 = XT(K)
          X3 = XT(K+1)
          Z1 = X-X1
          Z2 = X2-X
          Z3 = X3-X1
          SINTPD = Y1 + (1.0+Z2/Z3)*Z1*(Y2-Y1)/(X2-X1) - (Y3-Y2)/(X3-X2)*
         1      Z2*Z1/Z3
          RETURN
          END
```

@PRT,SC_ PCF.TPZ

14. TPZ

```
#EI.MHDDUCT.TPZ
                  DOUBLE PRECISION
                 1FUNCTION TPZ(P, Z, ZZ, NZ, NT, NP, VT, VP)
                  IMPLICIT DOUBLE PRECISION (A-H, O-Z)
                  DIMENSION ZZ(NZ, 1), VT(1), VP(1)
      C                     INTERPOLATE FOR T(P, Z) -- SIMILAR TO PZT.
                  DIMENSION VZT(30)
                  DO 10 I=1,NT
           10        VZT(I) = ZTP(ZZ, VT(I), P, NZ, NT, NP, VT, VP)
                  IF (VZT(1) .GT. VZT(NT))    GO TO 15
                     TPZ = SINTPA(VZT, VT, Z, NT)
                     RETURN
           15 CONTINUE
                     TPZ = SINTPD(VZT, VT, Z, NT)
                     RETURN
                  END

#PRT.SC   PCF.TZZ
```

15. DBLFF

```
EI•MHDDUCT•DBLFFF
              DOUBLE PRECISION
             1FUNCTION DBLFFF(X, XA, YA)
              IMPLICIT DOUBLE PRECISION (A-H, O-Z)
              DIMENSION XA(1), YA(1)
      C FUNCTION ROUTINE FOR SINGLE INTERPOLATION USED BY DINTRP
              X1 = XA(1)
              X2 = XA(2)
              X3 = XA(3)
              T1 = X - X1
              T2 = X - X2
              T3 = X - X3
              T = YA(1)*T2*T3/((X1-X2)*(X1-X3)) + YA(2)*T1*T3/((X2-X1)*(X2-X3))
             1    + YA(3)*T1*T2/((X3-X1)*(X3-X2))
              DBLFFF = T
              RETURN
              END

@PRT,SC  PCF.DUHDUCT
```

16. PWRE

```
EI*MHODUCT.PWRE
      DOUBLE PRECISION FUNCTION PWRE(PSTACK)
      IMPLICIT DOUBLE PRECISION (A-H, O-Z)
      COMMON /COMT2/ HAT(31), HWT(31),VTHWT(31),RWA,RWAP1,H2,HOXT(31),
     1        VBH(50),RMOX,RMAW,RMRP,VBT(50),NBT
      COMMON /REPR/HSTACK,TSTACK,PP(10),HRE
      DIMENSION GAM(15),TPROD(15)
C     GAM IS THE RATIO OF SPECIFIC HEATS TAKEN FROM THE 2502 OUTPUT
C     THE LISTED VALUES CORRESPOND TO ILLINOIS #6 COAL WITH 2.05% MOIST.
C     AND .7% SEED AS DRY K2CO3 FOR 333.33 -1111.1 DEG K AT ONE ATM
      DATA GAM /   1.3619D0,  1.3545D0,  1.3473D0,  1.3398D0,  1.3330D0,
     1  1.3260D0,  1.3190D0,  1.3116D0,  1.3056D0,  1.3001D0,  1.2950D0,
     2  1.2904D0,  1.2862D0,  1.2825D0,  1.2792D0
      ETAC = 0.9
      I = 0
      TRE = 500.
      TEMP = 333.333
      DO 10 I=1,15
      TPROD(I)=TEMP
   10 TEMP = TEMP + 55.555
   70 TA =(TSTACK + TRE)/2.
      TA2=TRE
      GP =SINTPA(TPROD,GAM,TA,15)
      TPI =TSTACK*(PP(2)/PSTACK)**((GP-1.)/GP)
      TRE = TSTACK + (TPI-TSTACK)/ETAC
      IF (ABS(TRE-TA2) .LE. .1) GO TO 100
      I= I+1
      IF (I .LT. 20) GO TO 70
   50 FORMAT(1H0,'NO SOLUTION' )
   80 PRINT 50
      GO TO 110
  100 HRE = SINTPA(  VBT,  VBH,TRE,NBT)
      PWRE = HRE -HSTACK
  110 RETURN
      END

PRT.SC  PCF.PZT
```

17. PZT

```
@EL*MHDDUCT.PZT
              DOUBLE PRECISION
             1FUNCTION PZT(Z, ZZ, T, NZ, NT, NP, VT, VP)
              IMPLICIT DOUBLE PRECISION (A-H, O-Z)
              DIMENSION ZZ(NZ, 1), VT(1), VP(1)
C                         INTERPOLATE FOR  P(T, Z)  BY FIRST OBTAINING
C                         VECTOR VZP(I) = ZZ(VP(I), T). THEN OBTAIN P FROM
C                         VZP(*), VP(*)
              DIMENSION  VZP(50)
              DO 10 J=1,NP
   10              VZP(J) = ZTP(ZZ, T, VP(J), NZ, NT, NP, VT, VP)
C                         CHECK FOR ORDERING OF VZP(*), ASCENDING OR DESCENDING,
C                         AND USE BINARY SEARCH TO LOCATE INTERPOLATION PTS.
              IF (VZP(1) .GT. VZP(NP))   GO TO 15
                   PZT = SINTPA(VZP, VP, Z, NP)
                   RETURN
   15         CONTINUE
                   PZT = SINTPD(VZP, VP, Z, NP)
                   RETURN
              END

@PRT,SC  PCF.SETUP
```

18. TZZ

```
EI*MHDDUCT.TZZ
               DOUBLE PRECISION
              1FUNCTION TZZ(S, ZS, H, ZH, NZ, NT, NP, VT, VP)
               IMPLICIT DOUBLE PRECISION (A-H, O-Z)
               DIMENSION ZS(NZ, 1), ZH(NZ, 1), VT(1), VP(1)
C                         INTERPOLATE FOR T(S, H).
C                        DEFINE G(T) = ZH(T, S) - H.  FIND G(T) = 0 BY SECANT
C                               METHOD.
               T  = VT(NT)
               G  = ZTZ(ZH, T, S, ZS, NZ, NT, NP, VT, VP) - H
               TO = VT(1)
               GO = ZTZ(ZH, TO, S, ZS, NZ, NT, NP, VT, VP) - H
             5 DG = G - GO
               DT = T - TO
               IF (DABS(DG*DT) .LE. 0.5D0)  GO TO 25
                    GI = GO
                    GO = G
                    TI = TO
                    TO = T
                    T = TI - GI*DT/DG
                    G = ZTZ(ZH, T, S, ZS, NZ, NT, NP, VT, VP) - H
                    GO TO 5
            25 TZZ = T
               RETURN
               END
```

PRT,SC PCF.ZPZ

19. ZPZ

WEI•MHDDUCT•ZPZ

```
                    DOUBLE PRECISION
                   1FUNCTION ZPZ(ZZ1, P, Z2, ZZ2, NZ, NT, NP, VT, VP)
                    IMPLICIT DOUBLE PRECISION (A-H, O-Z)
                    DIMENSION ZZ1(NZ, 1), ZZ2(NZ, 1), VT(1), VP(1)
      C                         INTERPOLATE FOR ZZ1(P, Z2)
                    T = TPZ(P, Z2, ZZ2, NZ, NT, NP, VT, VP)
                    ZPZ = ZTP(ZZ1, T, P, NZ, NT, NP, VT, VP)
                    RETURN
                    END
```

@PRT,SC PCF.ZTP

20. ZTP

```
EI.HHDDUCT.ZTP
              DOUBLE PRECISION
             IFUNCTION ZTP(ZT, X, Y, NZDIM, NX, NY, XT, YT)
              IMPLICIT DOUBLE PRECISION (A-H, O-Z)
              DIMENSION XT(1), YT(1), ZT(NZDIM,1)
      C DOUBLE INTERPOLATION USING LAGRANGE THREE-POINT METHOD
      C       FUNCTION DBLFFF IS REQUIRED
      C        ** PARAMETER **
      C           ZT = TABLE OF VALUES OF F(X,Y), ZT(I,J).
      C           XT = TABLE OF VALUES OF X.
      C           YT = TABLE OF VALUES OF Y.
      C           NX = NUMBER OF VALUES IN XT, ALSO NUMBER OF ROWS OF ZT.
      C           NY = NUMBER OF VALUES IN YT, ALSO NUMBER OF COLUMNS IN ZT.
      C           X,Y =  POINT FOR WHICH VALUE OF F IS DESIRED.
      C           NXDIM = ACTUAL ROW DIMENSION OF ZT.
      C       **NOTE.  EXTRAPOLATION USED IF EITHER X OR Y OUT OF RANGE.
              DIMENSION ZA(3), XA(3), YA(3), TZA(3)
              IK = NX-1
              JK = NY-1
              IDEX = 2
      C LOCATE POSITION OF X IN XT
              IF (XT(2).GT.X) GO TO 1000
              IDEX = IK
              IF (XT(IK).LE.X) GO TO 1000
              I = 1
          100 K = IDEX-I
              IF (K.LE.1) GO TO 1000
              J = (IDEX+I)/2
              IF (XT(J)-X) 200,300,300
          200 I = J
              GO TO 100
          300 IDEX = J
              GO TO 100
         1000 CONTINUE
      C LOCATE POSITION OF Y IN YT
              JDEX = 2
              IF (YT(2).GT.Y) GO TO 2000
              JDEX = JK
              IF (YT(JK).LE.Y) GO TO 2000
              I = 1
         1100 K = JDEX-I
              IF (K.LE.1) GO TO 2000
              J = (JDEX+I)/2
              IF (YT(J)-Y) 1200,1300,1300
         1200 I = J
              GO TO 1100
         1300 JDEX = J
              GO TO 1100
         2000 CONTINUE
              JDM1 = JDEX-1
              JDP1 = JDEX+1
              YA(1) = YT(JDM1)
              YA(2) = YT(JDEX)
              YA(3) = YT(JDP1)
      C DETERMINE ZT FOR Y AT XT(I), I=IDEX-1, IDEX, IDEX+1
              I = IDEX-1
              DO 3000 J=1,3
              ZA(1) = ZT(I,JDM1)
              ZA(2) = ZT(I,JDEX)
              ZA(3) = ZT(I,JDP1)
              TZA(J) = DBLFFF(Y, YA, ZA)
         3000 I = I+1
      C INTERPOLATE ABOVE RESULTS FOR VALUE AT X
              XA(1) = XT(IDEX-1)
              XA(2) = XT(IDEX)
              XA(3) = XT(IDEX+1)
              ZTP = DBLFFF(X, XA, TZA)
              RETURN
              END
```

@PRT.SC PCF.ZTZ

21. ZTZ

```
1*MHDDUCT.ZTZ
              DOUBLE PRECISION
             1FUNCTION ZTZ(ZZ1, T, ZZ, ZZ2, NZ, NT, NP, VT, VP)
              IMPLICIT DOUBLE PRECISION (A-H, O-Z)
              DIMENSION ZZ1(NZ, 1), ZZ2(NZ, 1), VT(1), VP(1)
      C                  INTERPOLATE FOR ZZ1(T, ZZ),   (USED BY TZZ)
              P = PZT(ZZ, ZZ2, T, NZ, NT, NP, VT, VP)
              ZTZ = ZTP(ZZ1, T, P, NZ, NT, NP, VT, VP)
              RETURN
              END

FIN
```

Appendix A 9.8

CARBONIZER AND SEPARATELY FIRED PREHEATER COMBUSTION CALCULATIONS

A 9.8.1 Carbonizer Calculations

The properties of the gaseous and char products of the carbonization process were determined from information supplied for the input coal, air requirements, and output weight fractions. The data sheets as supplied are contained in Tables A 9.8.1, A 9.8.2, and A 9.8.3. The composition of the tar are listed in Table A 9.8.4. The air required in the carbonizer process are shown in Table A 9.8.5.

Schematics of the carbonizer process using these data are seen in Figures A 9.8.1, A 9.8.2, and A 9.8.3. The air used in the carbonizer is 76.17% nitrogen, 23.19% oxygen, and 0.639% moisture by weight. The only unknown in the process was the composition of the char. This was found by performing a mass balance on the particular elements, knowing the fuel gas and tar compositions, weight fractions, and water weight fractions; and determining the char composition by assuming it was the residual of the process. Using this method, the composition of the various chars were calculated and are summarized in Tables A 9.8.6, A 9.8.7, and A 9.8.8 for the respective coals.

A 9.8.2 Preheater Combustion Study

It was specified that the flame temperature was not to exceed 2256°K (3600°F) in the gapor burner of the indirect fired preheater. With the lower rank coals this presented no problem because the gapor has a low heating value and a high water content. However, the flame temperature of the gapor from the carbonization of the Illinois No. 6 coals was in excess of this limitation and required that the flame temperature be

tailored by using exhaust gas recirculation. The exhaust products leaving the preheater on the gas side were chosen because if a gas was used whose temperature was below the preheater exit temperature, available heat would have to be supplied to the lower temperature stream. This presents problems in that no standard air-moving system can be utilized to recirculate the exhaust at the required temperature. As a result, it was decided to use the gapor combustion air stream to induce the recycled products.

In order to determine the heat available from the gapor combustion products for the main combustion air, it was necessary to know the thermodynamic properties of the gapor combustion products. This information was calculated by running a properties computer program (MHD 2502) for the composition of the carbonizer gapor (fuel gas, water, and tar) at an air equivalence ratio of 1.05 (5% excess air) over the temperature range of 1200 to 2700°K (1700 to 4400°F). The preheater combustion product leaving temperature was taken to be approximately 110°K (198°F) above the air inlet temperature. The heat required to get the combustion products to this temperature (QADD) was read from the aforementioned computer program. QADD is determined for the combustion of the specific gapor and air, both being at 288°K (59°F). Therefore, the sensible heat of the gapor and combustion air must be added to QADD to get the total heat available by cooling the gapor combustion products to the preheater exit temperature.

The ratio of the gapor mass flow to the main combustor oxidant mass flow (wet air plus recycled products) is determined from the ratio of the combustible mass flow to the main combustor oxidant mass flow and the ratio of gapor to the combustible mass flow (Tables A 9.8.1, A 9.8.2, and A 9.8.3). The ratios are calculated by the CHRDUC2 version of the duct program (Appendix A 9.7).

An energy balance is then simply done by

$$(MQ_{ADD})_G + (MC_p\ \Delta T)_{ca} + (MC_p\ \Delta T)_g = (MC_p\ \Delta T)_a \qquad \text{(A 9.8.1)}$$

where M is the mass flow rate, C_p is the specific heat at constant pressure, and ΔT is the change in temperature. On the gas side (denoted by g), $(MC_p\Delta T)_g$ and $(MC_p\Delta T)_{ca}$ are the sensible heats of the gapor combustion air, respectively, and $\dot{M}Q_{ADD}$ is the heat released from combustion of the gapor at $\phi = 1.05$ with the combustion products at the preheater exhaust temperature. On the air side (denoted by subscript a) M is the mass flow of the wet air and recycled products from the duct program. C_p is found by multiplying the specific heat of the air and recycled products by their respective mass ratios (a weighted average C_p is obtained). Knowing all other parameters, the air-side temperature rise was found. An iterative process was used which varied the amount of recycled products and kept the MHD combustor exit temperature a constant until the two temperatures were within 5°K (9°F).

Table A 9.8.1 - Properties of Low-Temperature Carbonization of Illinois No. 6 Bituminous Coal (Dry)

Temperature of Products - 800°F

Char

Weight fraction - 68%
HHV - 11,900 Btu/lb

Tar

Weight fraction - 9.4%
HHV vapor - 16,200 Btu/lb

Fuel Gas

Weight fraction - 37.55%
HHV - 3873 Btu/lb
Enthalpy*- 305.7 Btu/lb
Composition - mole fraction

CO_2	-	0.1635
O_2	-	0.0060
CO	-	0.0217
H_2	-	0.0554
C_2H_6	-	0.0425
CH_4	-	0.1724
N_2	-	0.5209
H_2S	-	0.0177

Light Oil

Weight fraction - 0.7%
HHV vapor - 17,000 Btu/lb

Ammonia

Weight fraction - 0.15%
HHV vapor - 9500 Btu/lb

Water - 4.3%

---

* Product enthalpy based on a reference temperature of 400°R

Table A 9.8.2 - Properties of Products of Low-Temperature Carbonization of North Dakota Lignite

Temperature of Products - 900°F

Moisture content of lignite as fired	27	18
lb air/lb coal as fired	0.67	0.59
Char		
Weight fraction	0.2597	0.3129
HHV - Btu/lb	11,955	11,995
Tar and L.O.		
Weight fraction	0.0262	0.0307
HHV vapor - Btu/lb	16,300	16,300
Fuel Gas		
Weight fraction	0.4955	0.472[illegible]
HHV - Btu/lb	438.8	521.[illegible]
Enthalpy* - Btu/lb	245.1	246.[illegible]
Composition - mole fraction		
CO_2 -	0.25[illegible]6	0.2679
CO -	0.0119	0.0144
H_2 -	0.0109	0.0132
CH_4 -	0.0220	0.0267
C_2H_4 -	0.0023	0.0028
H_2S -	0.0050	0.0050
N_2 -	0.6973	0.6700
Water		
Weight fraction	0.2186	0.1852

*Product enthalpy based on a reference temperature of 400°R

Table A 9.8.3 - Properties of Products of Low-Temperature Carbonization of Montana Subbituminous Coal

Temperature of Products - 900°F

Moisture content of coal as fired	20	16
lb air/lb coal as fired	0.56	0.52
Char		
Weight fraction	0.3395	0.3687
HHV - Btu/lb	12,230	12,240
Tar		
Weight fraction	0.0485	0.0517
HHV vapor - Btu/lb	16,200	16,200
Fuel Gas		
Weight fraction	0.4410	0.4257
HHV - Btu/lb	789.2	869.3
Enthalpy*- Btu/lb	250.7	252.1
Composition - mole fraction		
CO_2 -	0.2304	0.2314
CO -	0.0182	0.0199
H_2 -	0.0105	0.0113
CH_4 -	0.0433	0.0481
C_2H_4 = 0.043 -	0.0049	0.0054
H_2S -	0.0056	0.0055
N_2 -	0.6872	0.6785
Light Oil		
Weight fraction	0.0070	0.0070
HHV vapor	17,000	17,000
Water		
Weight fraction	0.1640	0.1460

* Product enthalpy based on a reference temperature of 400°R

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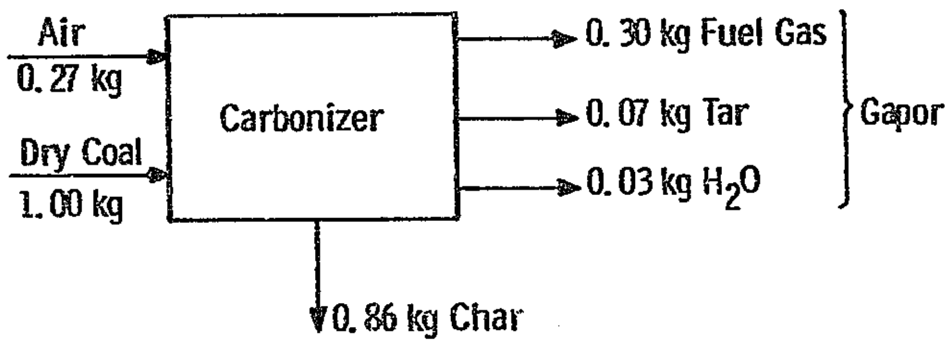

Fig. A 9.8.1—Schematic of carbonizer operating on Illinois No. 6 coal

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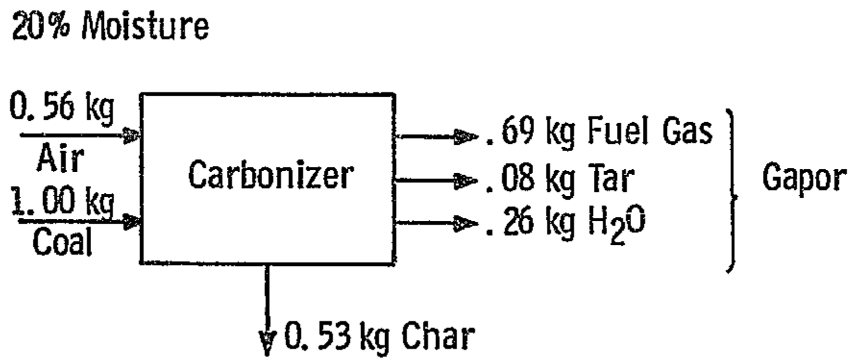

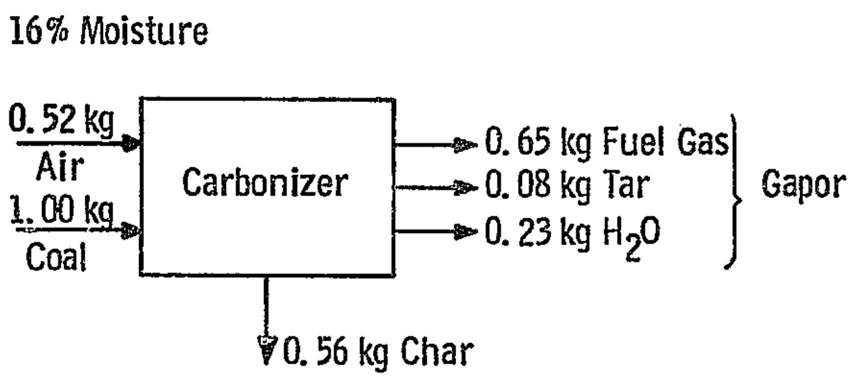

Fig. A 9.8.2— Schematic of carbonizer operating on Montana sub-bituminous coal

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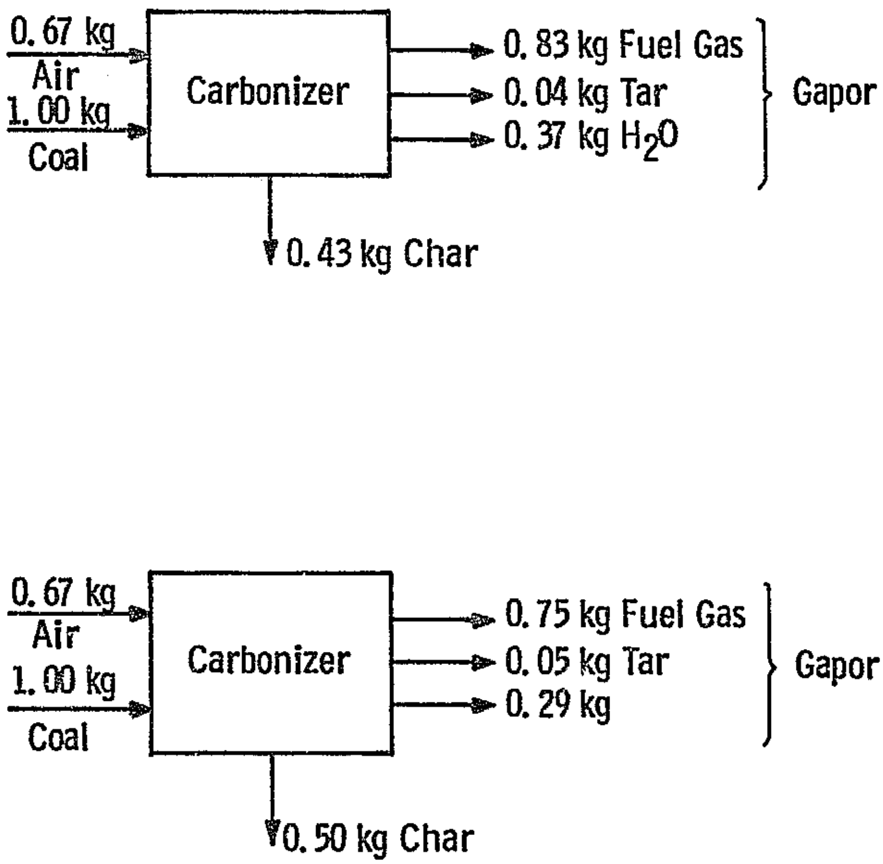

Fig. A 9.8.3—Schematic of carbonizer operating on North Dakota Lignite Coal

Table A 9.8.4 - Ultimate Analysis and Calorific Value of Dry Tars

		Low-Temperature Tars
C	%	82 - 8[illegible]
H	%	8 - 8.5
N	%	0.5 - 0.7
S	%	0.7 - 0.9
Ash	%	Negligible
O (by difference)	%	7.0 - 9.0
Calorific Value (Btu/lb)		
Gross		16,500 - 16,800
Net		15,700 - 16,000

(Data Supplied by the Coal Tar Research Association)

Table A 9.8.5 - Air Requirements for SRI Carbonization Process*

Coal	Mont.	Mont.	N. Dak.	N. Dak.	Ill.
Moisture - %	20	16	27	18	0
Temperature	900	900	900	900	800
Air lb/lb coal	0.56	0.52	0.67	0.59	0.27

* Stanford Research Institute

Table A 9.8.6 - Composition of the Char Produced by the Carbonization of Illinois No. 6 Coal (Predried)

Constituent	Wt. %
C	65.4
H	3.6
O	7.2
N	6.6
S	4.5
Ash	12.7

Table A 9.8.7 - Composition of the Char Produced by the Carbonization of Montana Subbituminous Coal

	Constituent	Wt. %
20% Moisture	C	76.7
	H	2.4
	O	4.3
	N	1.2
	S	0.8
	Ash	14.6
16% Moisture	C	76.9
	H	2.6
	O	3.9
	N	1.4
	S	0.8
	Ash	14.4

Table A 9.8.8 - Composition of the Char Produced by the Carbonization of North Dakota Lignite Coal

	Constituent	Wt. %
27% Moisture	C	78.8
	H	3.4
	O	0.0
	N	0.8
	S	0.8
	Ash	16.2
18% Moisture	C	78.0
	H	3.1
	O	0.0
	N	2.5
	S	0.9
	Ash	15.5

Appendix A 9.9

SUPERCONDUCTING MAGNET DESIGN FOR OPEN-CYCLE MHD GENERATORS

A 9.9.1 Introduction

The conceptual design of the superconducting magnet system for application to open-cycle MHD generators has been sufficiently developed for two generators (2000 and 600 MWe) to provide confidence in the design approach taken. The devices used and the conclusions drawn are described in this section.

The design of a superconducting magnet for MHD application does not require any new technological developments with respect to the superconductor. Preliminary examinations indicated that presently available multifilament niobium-titanium superconductors could be used in magnet systems with peak fields at the conductor up to 7.5 T at 4.2°K (-452.13°F). For magnet systems with peak fields at the conductor between 7.5 and 10.0 T and operating at 4.2°K (-452.13°F) projections (based on currently funded conductor development programs for fusion and ac generator applications) indicate that multifilamentary niobium-titanium superconductors will be commercially available in 1990. These limitations are based on a judgment concerning the degree of margin required in a large magnet system.

The major consideration in superconducting magnet design for MHD application is selecting a magnet configuration which can be confined by an economical mechanical structure. For large MHD systems two magnet configurations are often considered. For mechanical design, the rectangular type of magnet configuration seems preferable to the circular with the same peak fields in the windings. Considering the desired field uniformity, however, and the magnitude of the magnetic field on axis of the MHD duct, the circular cross-sectional winding

configuration is more desirable. This winding can be shaped to achieve the same peak field in thie winding as the field on the axis of the MHD duct, whereas with the rectangular magnet configuration the peak field at the winding is greater than the field on axis and is strongly dependent upon the size of the winding relative to the duct and the desired uniformity.

Accordingly, the base case designs described herein have been determined by utilizing the following design conditions:

- A niobium-titanium (Nb Ti) filamentary conductor would be employed for magnets with peak fields less than 7.5 T at the conductor.
- A niobium-tin (Nb_3Sn) filamentary conductor would be used for magnets with peak fields greater than 7.5 T at the conductor.
- The magnet winding would be force cooled.
- The mechanical structure would be designed to be self-supporting against magnetic forces.
- The system would be designed for an overall minimum cost.
- The magnet configuration would be selected in such a way that the variations of the field in the cross-sectional area of the duct would be less than 5% transverse and less than 8% parallel to the magnetic field.

A 9.9.2 Electrical Design

The minimum cost of superconducting winding is obtained when the maximum current density is employed consistent with adequate operational reliability. In the initial stages of this project two candidate magnet configurations were considered. The first was the rectangular type. Magnetic field profiles were determined by using the general magnetic field code MAFCO developed by Lawrence Radiation Laboratory (LRL) (Reference 9.38). Field uniformity requirements were

c5

specified to be less than 5% variation transverse and less than 8% variation parallel to the magnetic field.

The minimum cost rectangular cross-sectional field winding distribution that met the above uniformity limitations was sought To achieve the desired uniformity, the currents had to be distributed around the entire perimeter of the duct. Unfortunately, the design study indicated that in order to achieve the desired uniformity, the peak fields at the windings were more than 25% greater than the magnetic field on the axis of the duct.

The second magnet configuration considered was the circular saddle coil. This configuration is similar to the dipole configurations used in beam-line magnets for accelerators and readily lends itself to analytical investigation. The peak magnetic field in circular saddle coils can be designed to be the same as the base field, and the uniformity achieved within the base of the magnet depends solely upon how closely the actual winding configuration approximates the ideal solution.

Since the nominal field in the duct of an open-cycle MHD generator is 6 T, the preliminary investigation indicated that the rectangular magnet configuration would require niobium-tin filamentary conductors and that the circular saddle magnet configuration would require niobium-titanium filamentary conductors. Further investigation indicated that although the circular saddle configuration required less conductor, it would require more structure than the rectangular magnet configuration. The circular saddle was selected because of the greater degree of margin offered. The low magnetic field at the conductor affords higher reliability.

A 9.9.2.1 Conductor Design

The environment of an MHD magnet system does not require the selection of sophisticated superconductors, primarily because there are no ac or transient magnetic fields present. Furthermore, the magnetic

field distribution cannot accommodate graded winding designs because of the conical dipolar configuration.

Since the nominal magnetic field seen by the MHD duct for the open-cycle concept is 6 T and the estimated peak field at the superconducting windings (in the end turns at the entrance of the duct) is less than 7.5 T, filamentary niobium-titanium conductors have been selected for the base case designs.

The most important aspect of the conductor design is the selection of winding current density such that the winding will not quench during conventional operation. Figure A 9.9.1 illustrates a normalized plot of the critical current density for niobium-titanium wires as a function of the peak field on the wire [where jc(5T) $\sim$ 2×10^9 A/m^2]. The operating current of the magnet system was arbitrarily set at 5000 A. A conductor with filaments of 100 μm (0.394 mills) or less, must have approximately 1 twist/2.54 cm (1 in) to inhibit the possibility of flux jump instability in the wire when the magnet is being either charged or discharged. Using a conductor with an aspect rate of 2:1, a winding packing factor, λ, of 0.1 was established as sufficient spacing to accommodate liquid helium cooling ducts, headers, and the necessary distributed support structure within the winding. To ensure stable operation a 3:1 copper to superconductor ratio was selected which does not have enough copper to cryostabilize the winding but should have enough to dynamically stabilize it (Reference 9.39). Although the MHD magnet is a dc device, provision for operational margin must be provided. If one has temperature excursions of 0.1 to 0.2°K (0.18 to 0.36°F) from the nominal 4.2°K (-452.13°F) in the windings, an operational current density j equal to 0.5 jc at 4.2°K (-452.13°F) affords a reasonable compromise between high current density and thermal margin and was selected as the design point for the peak field region in the winding. Accordingly, a winding current density, λJ, of 1.4×10^8 A/m^2 was selected for the electrical design of the open-cycle MHD base case magnet designs.

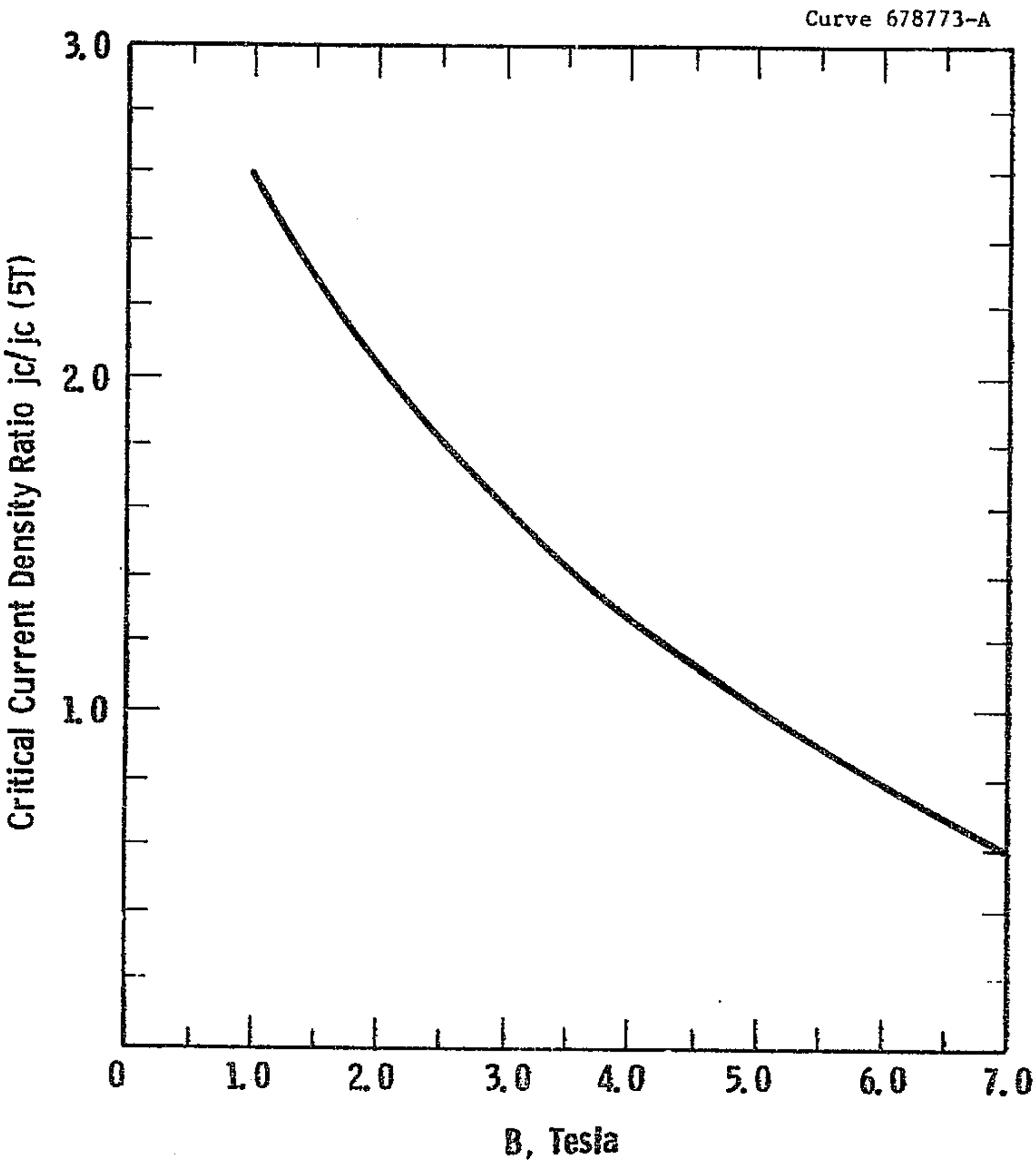

Fig. A 9.9.1—Normalized J_c-B curve for NbTi superconductors

A 9.9.2.2 Magnetic Field Analysis

The electrical design for the open-cycle MHD generators is based upon a dipole current distribution configuration. The essential features and parameters of the dipole design are shown in Figure A 9.9.2 which depicts a cross section of the MHD generator. The ideal dipole construction consists of two overlapping circles having diameters D and a separation between circle centers of d. The winding areas are defined by the nonoverlapping regions of the offset circles. In the ideal case the magnetic field in the overlap region is constant in magnitude and direction. The generator duct, which is an L by L square in cross section, is centered in this constant field region in such a way that the duct corners are a distance R from the inner surface of the windings. In general, the area of the duct and the magnetic field strength vary with distance down the generator axis.

The design equations used to calculate the dipole geometry and current distribution as a function of the distance down the duct assume a constant winding current density λJ (A/m^2) in the winding areas. For overlapping circles the magnetic field B inside the dipole windings is related to the separation of centers d by

$$d = 2B/\mu(\lambda J) \qquad \text{(A 9.9.1)}$$

Considering the geometry of Figure A 9.9.3 the diameter of the circles is given by

$$D = \sqrt{(L + \sqrt{2}\,R + d)^2 + (L + \sqrt{2}\,R)^2} \qquad \text{(A 9.9.2)}$$

and the winding area A of each current source is calculated from

$$A = \frac{d}{2}\sqrt{D^2 - d^2} + \frac{D^2}{2}\ \arcsin\left(\frac{d}{D}\right) \qquad \text{(A 9.9.3)}$$

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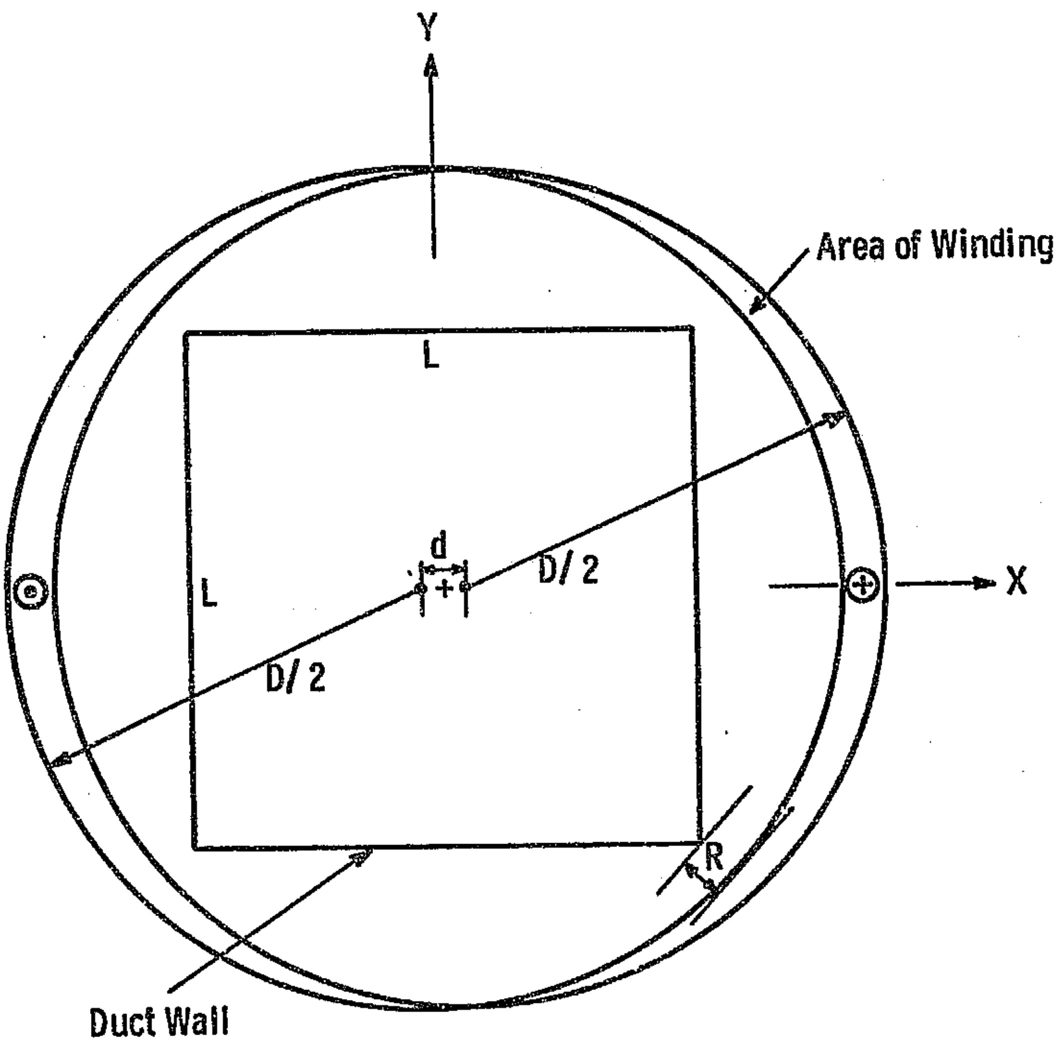

Fig. A 9. 9. 2—Dipole magnet design for the open-and closed-cycle, inert gas MHD generators

which for $d << D$ reduces to $A \cong \frac{dD}{2}$. The winding current is computed from

$$NI = A\ (\lambda J)$$

Once λJ and R are fixed and the magnetic induction, B, and duct size, L, are given as a function of the distance, Z, down the duct axis, then the dipole design is determined at any cross section of the generator by the equations above.

The electrical designs for Base Case 1, Point 3 and Base Case 2, Point 1 open-cycle generators are presented in Table A 9.9.1. The winding current and magnetic field distributions down the generator axis are shown in Figure A 9.9.3 for both designs. The peak winding currents for magnetic motive force, MMF, for these designs are 2.214×10^7 and 3.703×10^7 ampere-turns, respectively.

The above description assumes an ideal current distribution to be used in determining the cost of the superconducting magnet system. The actual magnet configuration requires a more detailed examination of the magnetic field distribution. The approach to be used in determining a detailed electrical design has been determined and is outlined herein.

The division of the ideal current distribution into discrete layers of conductors is made with the constraint that a certain field uniformity is desired within the base of the magnet. The uniformity can be accurately determined at each cross section by summing the induction produced by each shell of current. Figure A 9.9.4 illustrates a schematic of the shell configuration. The magnetic induction produced by the j-th shell can be approximately calculated from Reference 9.40.

Table A 9.9.1 - Dipole Magnet Design for the Open-Cycle, Inert Gas MHD Base Line Generators with r = 6 in.

Base Case 1, Point 3

I	Z,m	L,m	D,m	d,m	NI A-Turn	Mag. Ind. (B),T
1	0.000	1.007	1.832	0.1425	0.17480+08	6.00
2	0.081	1.010	1.837	0.1425	0.17520+08	6.00
3	0.809	1.044	1.885	0.1425	0.17983+08	6.00
4	1.543	1.080	1.936	0.1425	0.18472+08	6.00
5	2.286	1.122	1.994	0.1425	0.19029+08	6.00
6	3.043	1.169	2.061	0.1425	0.19668+08	6.00
7	3.822	1.221	2.135	0.1425	0.20375+08	6.00
8	4.629	1.282	2.221	0.1425	0.21191+08	6.00
9	5.475	1.352	2.320	0.1425	0.22142+08	6.00
10	6.442	1.437	2.431	0.1311	0.21350+08	5.52
11	7.670	1.538	2.567	0.1200	0.20623+08	5.05
12	9.283	1.665	2.738	0.1090	0.19998+08	4.59
13	12.018	1.869	3.016	0.0960	0.19391+08	4.04
14	13.264	1.954	3.134	0.0915	0.19201+08	3.85
15	14.744	2.054	3.272	0.0869	0.19056+08	3.66
16	16.534	2.170	3.432	0.0824	0.18954+08	3.47
17	16.647	2.177	3.442	0.0822	0.18953+08	3.46

Table A 9.9.1 (Cont'd) - Dipole Magnet Design for the Open Cycle, Inert Gas MHD Base Line Generators with r = 6 in.

Base Case 2, Point 1

1	Z,m	L,m	D,m	d,m	NI A-Turn	Mag. Ind. (B),T
1	0.000	1.831	2.996	0.1425	0.28600+08	6.00
2	0.085	1.836	3.004	0.1425	0.28674+08	6.00
3	0.864	1.898	3.092	0.1425	0.29514+08	6.00
4	1.667	1.964	3.185	0.1425	0.30404+08	6.00
5	2.500	2.039	3.291	0.1425	0.31418+08	6.00
6	3.369	2.123	3.410	0.1425	0.32555+08	6.00
7	4.282	2.219	3.545	0.1425	0.33841+08	6.00
8	5.251	2.327	3.697	0.1425	0.35299+08	6.00
9	6.290	2.455	3.879	0.1425	0.37029+08	6.00
10	7.509	2.605	4.082	0.1311	0.35860+08	5.52
11	9.103	2.788	4.333	0.1200	0.34824+08	5.05
12	11.264	3.015	4.646	0.1088	0.33865+08	4.58
13	15.089	3.379	5.152	0.0960	0.33126+08	4.04
14	16.888	3.533	5.366	0.0915	0.32881+08	3.85
15	19.076	3.709	5.612	0.0869	0.32687+08	3.66
16	21.787	3.916	5.901	0.0822	0.32493+08	3.46
17	22.067	3.936	5.929	0.0820	0.32555+08	3.45

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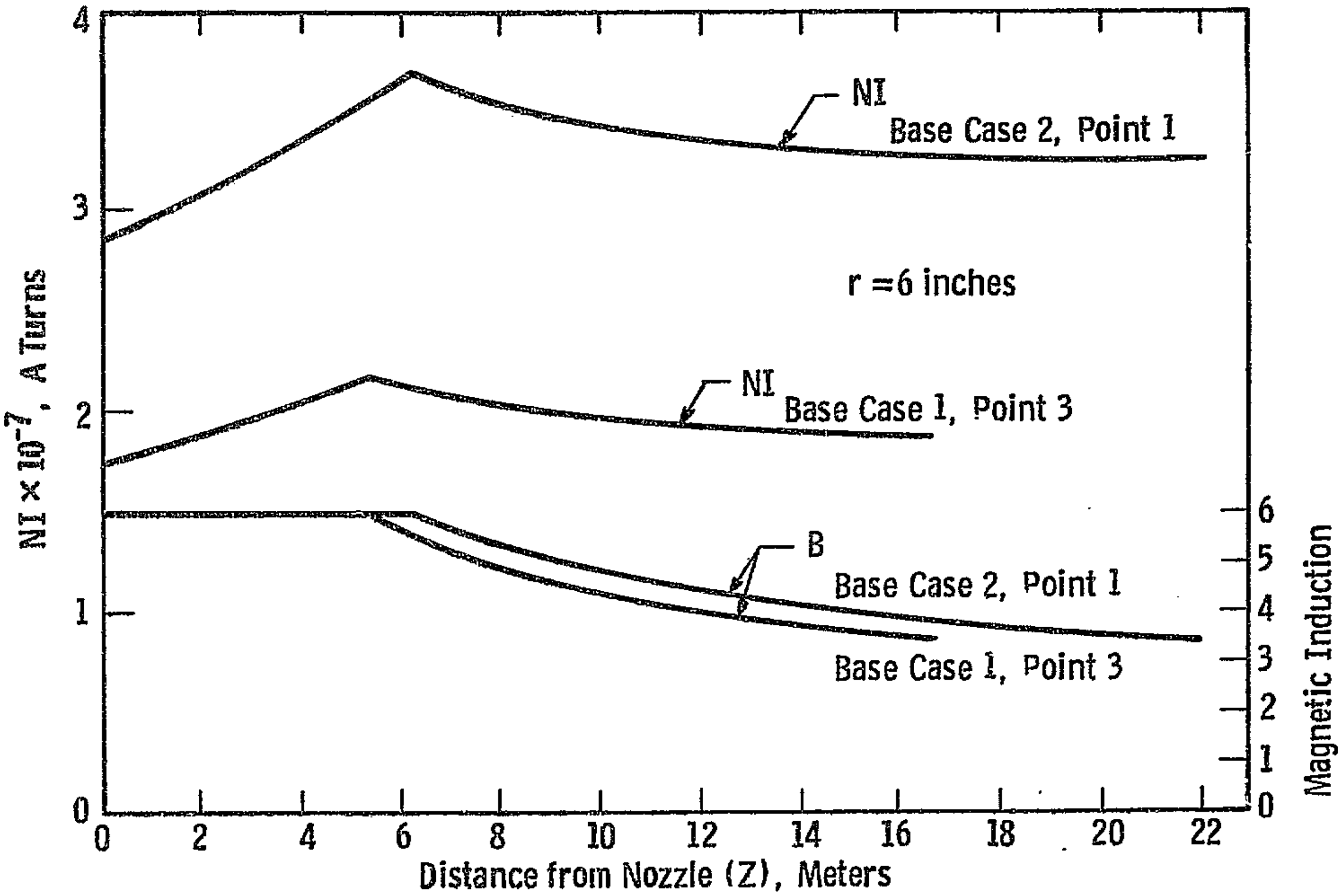

Fig. A 9.9.3—Winding current and magnetic field as a function of distance down the duct axis for direct-fired open-cycle MHD generator (Base Case 2)

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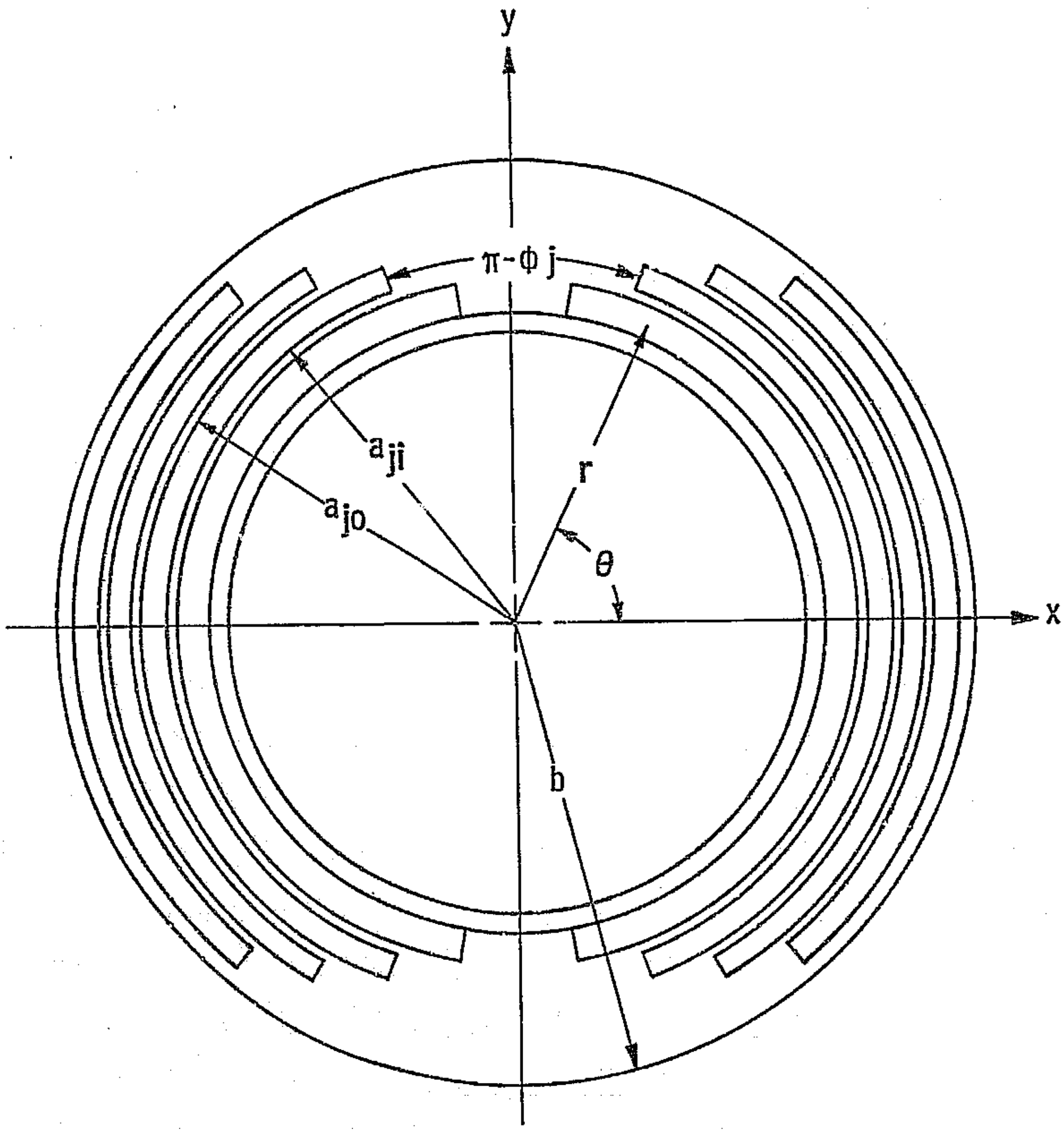

Fig. A 9.9.4—Shell-type dipole winding

$\underline{r < a_{ji}}$

$$H_r = - \sum_{n=odd} \frac{2 J_j \sin(\frac{n\phi_j}{2}) \sin (n\theta)}{n \pi (2-n)} \; r\left(\frac{r}{a_{jo}}\right)^{n-2} \left[1 - \left(\frac{a_{ji}}{a_{jo}}\right)^{2-n}\right] \qquad \text{A 9.9.5}$$

$$H_\theta = \sum_{n=odd} \frac{2 J_j \sin(\frac{n\phi_j}{2}) \cos (n\theta)}{n \pi (2-n)} \; r\left(\frac{r}{a_{jo}}\right)^{n-2} \left[1 - \left(\frac{a_{ji}}{a_{jo}}\right)^{2-n}\right] \qquad \text{A 9.9.6}$$

$\underline{a_{ji} < r < a_{jo}}$

$$H_r = - \sum_{n=odd} \frac{2 J_j \sin(\frac{n\phi_j}{2}) \sin (n\theta)}{n \pi (4-n^2)} \; r \left[- 2n - (2-n)\left(\frac{a_{ji}}{r}\right)^{n+2} + (2+n) \left(\frac{r}{a_{jo}}\right)^{n-2}\right] \qquad \text{A 9.9.7}$$

$$H_\theta = - \sum_{n=odd} \frac{2 J_j \sin(\frac{n\phi_j}{2}) \cos (n\theta)}{n \pi (4-n^2)} \; r \left[- 4 + (2-np)\left(\frac{a_{ji}}{r}\right)^{n + 2} + (2+n) \left(\frac{r}{a_{jo}}\right)^{n-2}\right] \qquad \text{A 9.9.8}$$

$\underline{a_{jo} < r}$

$$H_r = - \sum_{n=odd} \frac{2 J_j \sin(\frac{n\phi_j}{2}) \sin (n\theta)}{n \pi (2+n)} \; r \left(\frac{a_{jo}}{r}\right)^{n+2} \left[1 - \left(\frac{a_{ji}}{a_{jo}}\right)^{n+2}\right] \qquad \text{A 9.9.9}$$

$$H_\theta = - \sum_{n=odd} \frac{2 J_j \sin(\frac{n\phi_j}{2}) \cos (n\theta)}{n \pi (2+n)} \; r \left(\frac{a_{jo}}{r}\right)^{n+2} \left[1 - \left(\frac{a_{ji}}{a_{jo}}\right)^{n+2}\right] \qquad \text{A 9.9.10}$$

where H_r is the magnetic field in the radial direction at r and θ; H_θ is the magnetic field in the circumferential direction at r and θ, and J_j is the average current density in the j-th shell.

Finally, one can field map the complete circular saddle coil using a general magnetic field code such as MAFCO-W (Reference 9.41) which is an extended version of MAFCO developed by the University of Wisconsin. This code has the capability of determining the field at any point in space generated by arc segments and line segments of rectangular solid conductors carrying current.

A 9.9.2.3 Stored Energy and Inductance Considerations

The size of the superconducting magnet for the open-cycle MHD system is considerably larger than today's large bubble chamber magnets and will be larger than the magnets required for fusion applications. Extrapolation from existing large magnets to magnets required for MHD application should be done with care. Unfortunately, the technology for large magnet systems being evolved for the fusion program will not be fully applicable to MHD magnet development;and, therefore, before an attempt is made to make a large-scale superconducting magnet for MHD application, investigation of the behavior of an intermediate-sized MHD magnet system typical of the larger-scaled version should be considered to uncover any technological problems.

One fact that has been considered and is reported here is that the large size of this magnet implies that there is a large stored energy. This represents a two-fold problem. First, the energy stored per unit volume of conductor makes quench detection, protection, and prevention an important aspect of the electrical design of such a large magnet system. Second, the time required to excite the magnet to full field can be considerable, depending upon the design voltage stress that conductor insulation can withstand.

For the ideal winding distribution the inductance, L, was found to be approximately given by:

$$L, \text{henries} = \frac{\pi \mu_o N^2 \ell}{8} \qquad \text{A 9.9.11}$$

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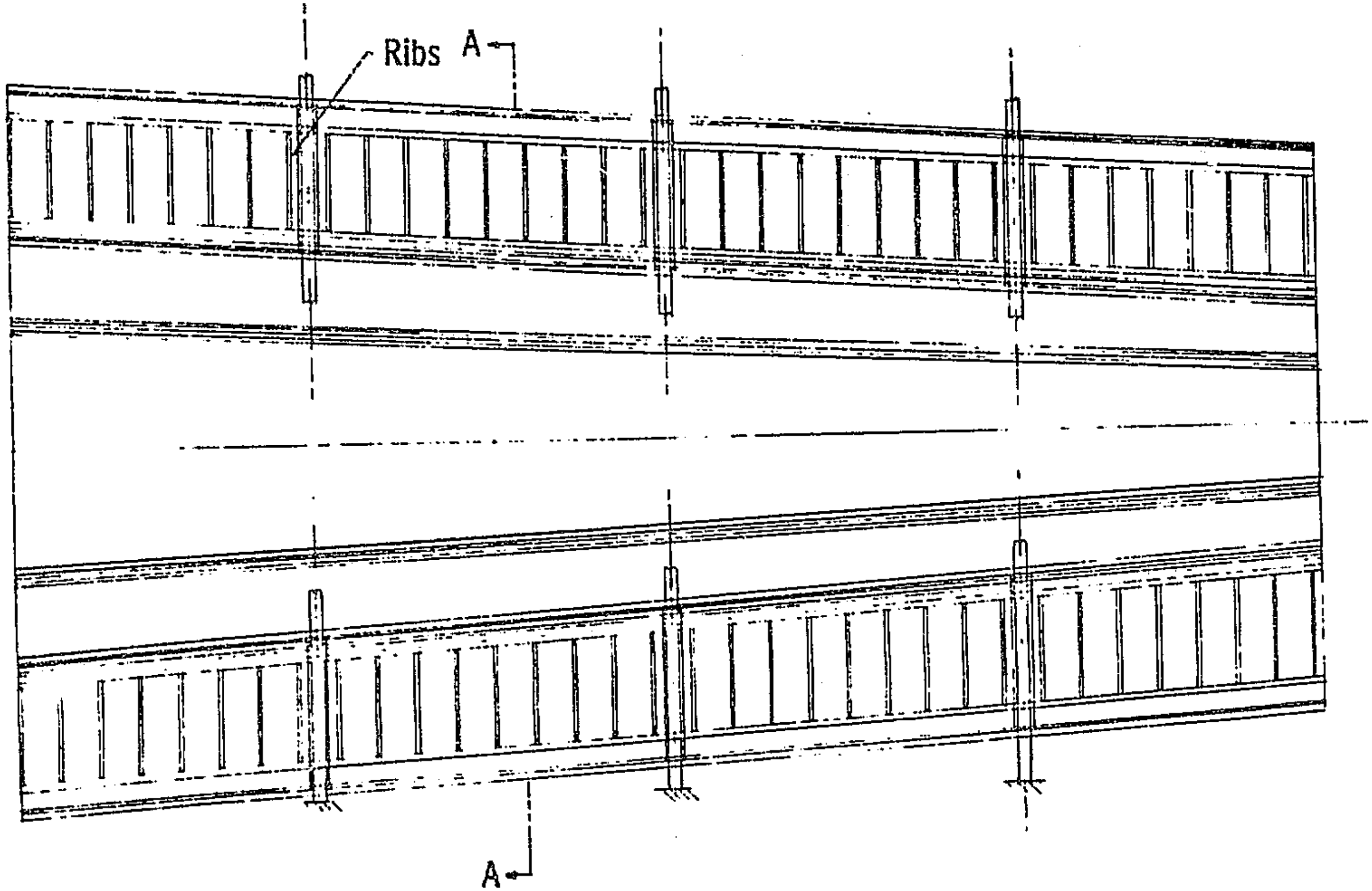

Fig. A 9.9.5 - Sectional view of MHD duct and support structure

where N is the number of turns, ℓ is the axial length in meters of the winding, and μ_o is permeability of free space, 4×10^{-7} T/(A/m). A more detailed examination for the real winding distribution will give an answer very close to this one.

Since the typical MHD magnets considered have very large stored energies, it is recommended that the winding be subdivided and powered by separate power supplies in order to ensure reliable operation and decrease the voltage seen during a quench.

A 9.9.3 Mechanical Analysis

The major structural problem in the design of the dewars is the provision of a lightweight wall structure (to minimize construction costs) that is compatible with the bending moments generated by the electromagnetic forces on the coils. Because of the need to operate the dewar at liquid helium temperature, it is impractical from a heat transfer standpoint to provide supporting struts to the warmer structural elements; and, therefore, the dewars were designed to be essentially self-supporting, except for the provision of support columns at the base to hold the weight of the structure.

For maximum economy and minimum weight, the dewar walls subjected to bending loads were designed in a plate-girder form. The spacing, height, and thickness of the webs were optimized with the thickness of the plates to produce a minimum weight structure consistent with the design stresses and buckling. In some cases, where other design considerations were dominant, a nonoptimum (in terms of weight) structure was used.

A 9.9.3.1 General Arrangement

The magnet dewar and vacuum chamber are frustums of right circular cones which have their major axes horizontal, as shown in Figure A 9.9.5. The superconducting magnet and its associated structure are supported against the gravitational force by many circular cross-sectional columns which are stage cooled with liquid nitrogen to minimize the conduction losses. Similarly, the

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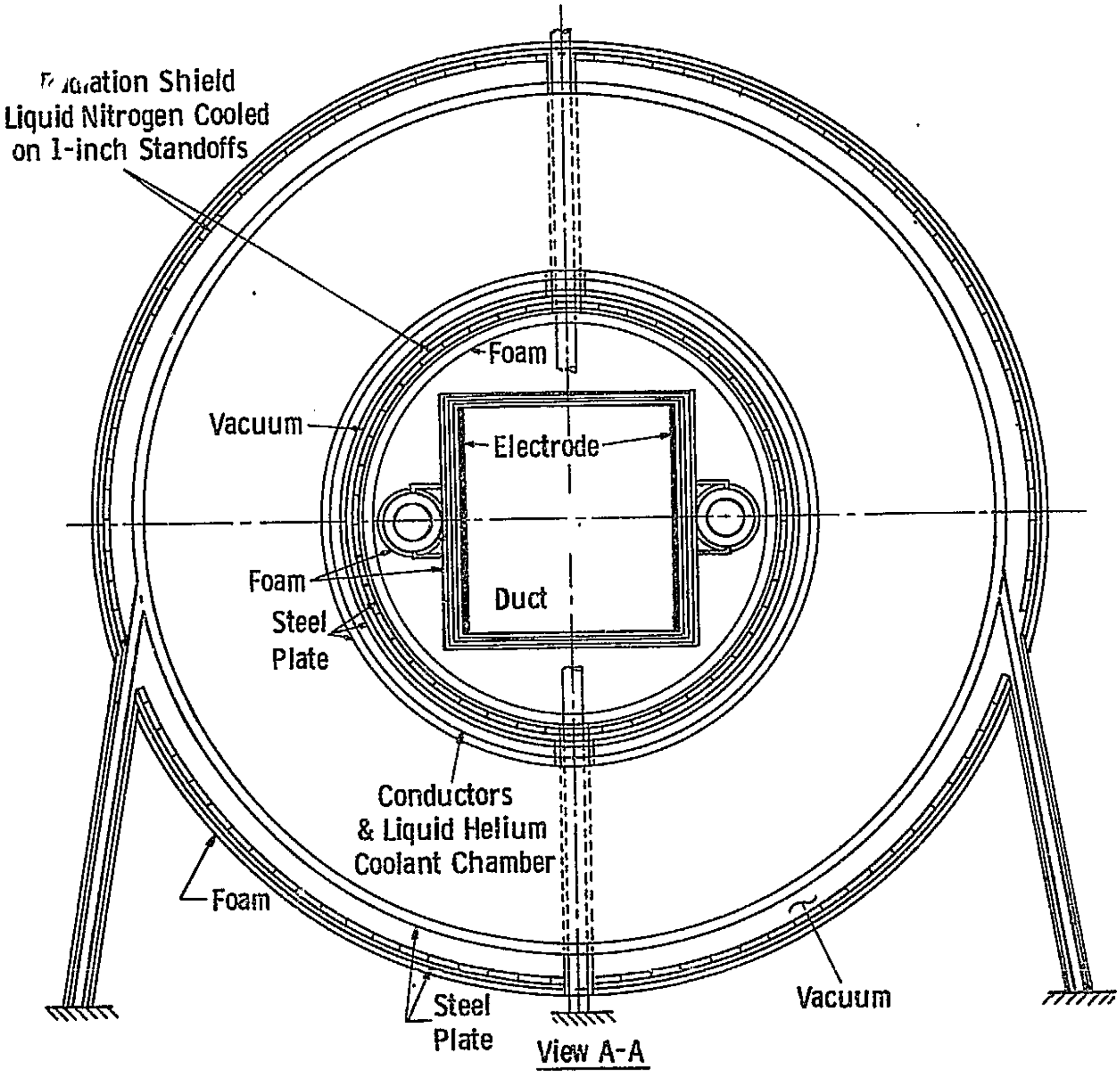

Fig. A 9.9.6—Cross sectional view of MHD duct and magnet enclosure

liquid nitrogen vacuum chamber is supported by concentric tubular columns. The inner vacuum chamber is attached to the outer vacuum chamber at the far ends of the magnet and at several locations along the length through vertical tubular struts. These arrangements are illustrated in Figure A 9.9.6. The magnet dewar is designed to restrain conductor motion under magnetic loading. Hence, the minimum cost design is the conical equivalent of I-beam construction for the outer dewar wall, as shown in Figures A 9.9.6 and A 9.9.7. In order to distribute the magnetic loading and restrain conductor motion, the space between the inner dewar wall and the outer dewar walls is fully occupied, as shown in Figure A 9.9.7, either by the superconductor, by the cooling ducts, or by a stainless steel filler. A study of using foam insulation over a liquid nitrogen shield versus radiating from room temperature indicated a factor of six reduction in the liquid nitrogen cooling requirements for the latter case. Foam insulation was, therefore, selected as the best means of enclosing the magnet structure.

This general arrangement was selected because during cool-down the 0.3% thermal contraction of the magnet and support structure does not subject any of the materials to high thermal stresses.

A 9.9.4 Heat Transfer Analysis

The requirements of maintaining a superconducting magnet at 4.2°K (-452°F) within a tolerance of 0.2°K (0.36°F), can be achieved with existing technology. Liquid helium and nitrogen refrigeration systems can be operated continuously for a year or more. In the past, the limiting factor in system endurance was clogging of the flow passages due to freezing out of impurities. The major impurity source was compressor oil. Presently, dry compressors, turboexpanders, and redundant compressors in a closed-loop cryogenic refrigeration system can be operated continuously for several years. Cryogenic refrigeration equipment with sufficient reliability to support the superconducting magnets for MHD generators are available.

A survey of the manufacturers of refrigeration equipment (Reference 9.42) was implemented several years ago, and the results have been applied in estimating the electrical requirements of the refrigeration

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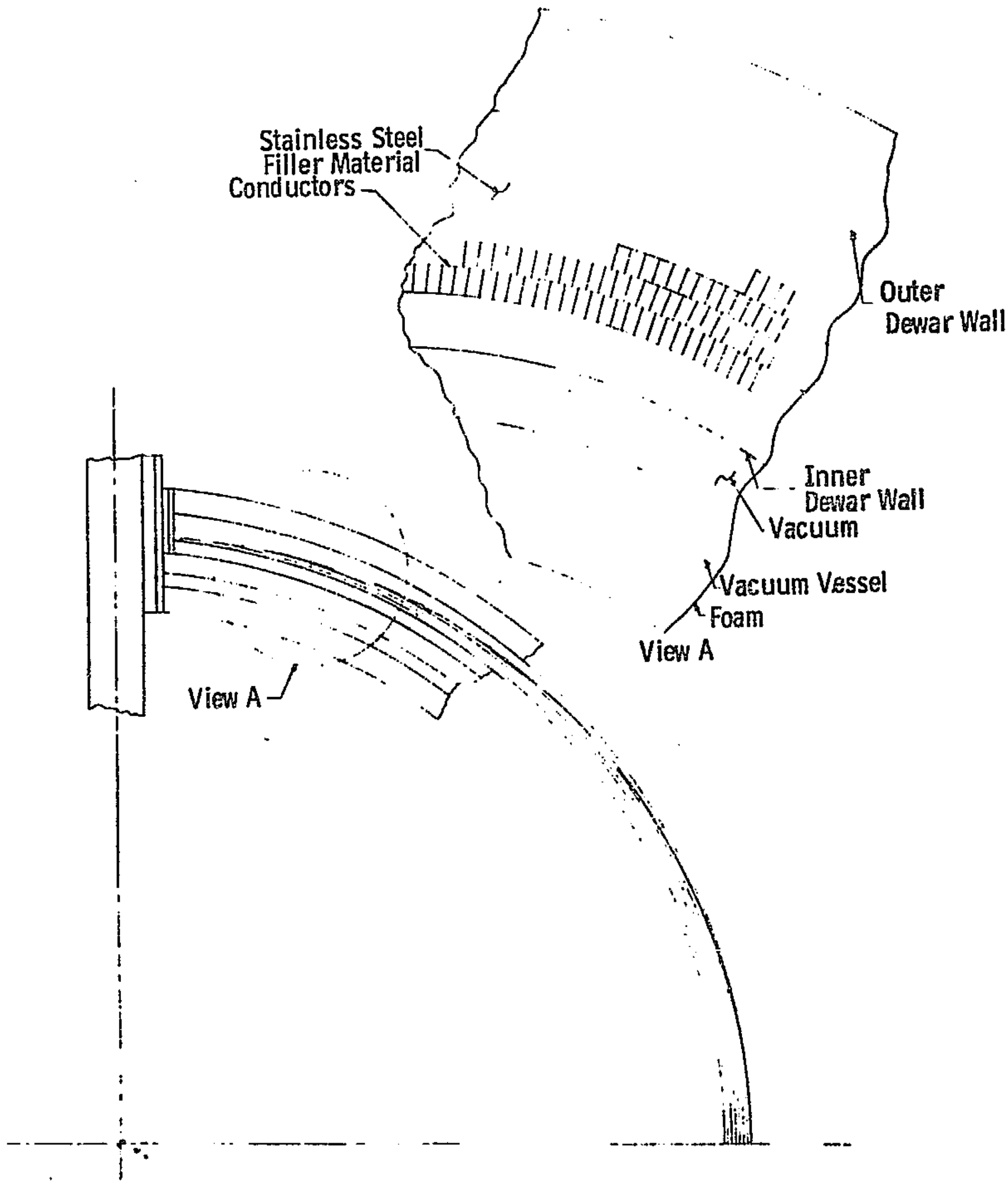

Fig. A 9.9.7 - Sectional view of winding distribution

equipment. No allowance has been made for technological improvements that may be made to increase the efficiency of these machines. It is expected, however, that the continued emphasis in utilizing superconducting magnets in power generation systems will encourage innovation in the heat exchanger design and improvements in the reliability.

Liquid helium refrigeration systems can be obtained with specific power requirements (watts of electrical power required to produce one watt refrigeration) as low as 300 W/W and helium liquefaction systems with as low as 900 W/W. These specific powers are generally obtained only in very large-capacity devices. For a total system load of 120 W, 600 W/W are required for refrigeration and 1800 W/W for liquefaction. For a total system load of 280 W, 550 W/W are required for refrigeration and 1650 W/W for liquefaction.

Liquid nitrogen refrigerator systems can be obtained with specific power requirements as low as 6.8 W/W. For a 20 kW load, a specific power requirement of 7 W/W is generally obtained, and for a 9 kW load a specific power of 8 W/W is usually required.

A cursory examination of the use of foam insulation around the periphery of the liquid nitrogen shield showed that the liquid nitrogen requirements for a radiation shield would be one-sixth those for a foam-insulated liquid nitrogen vacuum vessel. Since the most economical system for a magnet operating in a power system is usually the one with the highest efficiency, the lower-loss shielding system was selected. If accessibility is an important requirement, the foam insulation will not severely affect the overall plant efficiency.

Tables A 9.9.2 and A 9.9.3 present a summary of the cooling requirements for the open-cycle MHD magnets. For the 600 MVA design (like Base Case 1, Point 3) 400 m^2 (4306 ft^2) of surface are exposed to radiation, 12 electrical leads are recommended to power the magnet, and 10 steel 2.44 m (8 ft) columns long and totalling a cross-sectional area of 0.163 m^2 (1.75 ft^2) are recommended to support the 422 Mg stainless steel dewar. For the 2000 MVA design (like Base Case 2, Point 1), 1000 m^2 (10,764 ft^2) of surface is exposed to radiation, 20 electrical leads are recommended to power the magnet, and 36 steel columns

Table A 9.9.2 Summary of Cooling Requirements
2000 MVA Design

a) Helium Refrigeration

Load	Heat Input, W	Mass Flow lb/hr	Electrical Load, kW
Radiation	32	36	17.6
Electrical Leads	125	140	206.3
Support Structure	120	135	66.0
Totals	277	311	289.9

b) Nitrogen Refrigeration

Helium Refrigerator	1335	53	9.3
Radiation	19933	798	139.5
Conduction	975	40	6.8
Totals	22243	891	155.6

Table A 9.9.3 Summary of Cooling Requirements
600 MVA Design

a) Helium Refrigeration

Load	Heat Input, W	Mass Flow lb/hr	Electrical Load, kW
Radiation	12.8	14.4	7.7
Electrical Leads	75.0	84.0	135.0
Support Structure	33.3	37.5	20.0
Totals	121.1	135.9	162.7

b) Nitrogen Refrigeration

Helium Refrigerator	583	23.2	4.7
Radiation	7973	319.2	63.8
Conduction	271	11.1	2.2
Totals	8827	353.5	70.7

2.44 m (8 ft) long and a total cross-sectional area of 1.11 m^2 (12 ft^2) are required to support the 1.63 Gg (1800 tons) stainless steel dewar.

The 600 MVA design requires a 120 W helium refrigerator-liquefier which should occupy approximately 25.5 m^3 (900 ft^3) and weigh 8.165 Mg (9 tons). Furthermore, a 9 kW nitrogen refrigerator is required and will occupy 4.25 m^3 (150 ft^3) and weigh approximately 3 Mg (3.3 tons).

The 2000 MVA design requires a 280 W helium refrigerator-liquefier which should fill a volume of 51.0 m^3 (1800 ft^3) and weigh approximately 14.5 Mg (16 tons). Furthermore, a 22 kW nitrogen refrigerator will occupy 7.65 m^3 (270 ft^3) and weigh 5.443 Mg (6 tons).

The installed cost of the above refrigeration system is $300,000 for the 600 MVA system and $400,000 for the 2000 MVA system.

A 9.9.5 Summary

In the cost study performed on the open-cycle MHD magnet system the following parametric expression for the cost of the conductor was developed:

$$C = \frac{4.8 \, B_{av} \, Z}{\mu_o \lambda J_{av}} \left(\frac{L_1 + L_2}{\sqrt{2}} + 2R \right)$$

where
- C = cost of conductor, \$ x 10^{-6}
- B_{av} = average magnetic induction field, T
- Z = length of duct, m
- μ_o = $4\pi \times 10^{-7}$ = permeability of free space
- λJ_{av} = average winding current density, A/m^2
- L_1 = inlet duct width, m
- L_2 = exit duct width, m
- R = insulation thickness, m.

For the two baseline designs evolved the insulation thickness was made 15.24 cm (6 in).

Summaries of the results obtained on the two magnet designs are given in Table A 9.9.4. Cost breakdowns for these two cases are given in Tables A 9.9.5 and A 9.9.6.

The component costs, weights, and heat loads were linearly scaled from the detailed base cases presented, using the cost of the conductor for the prescribed duct geometries and magnetic field profiles as the fundamental scaling factor.

Table A 9.9.4 Summary of the Open-Cycle MHD Magnet Designs

	(like Base Case 1, Point 3)	(like Base Case 2, Point 1)
Nominal Plant Rating	600 MVA	2000 MVA
Inlet Cross-Sectional Area	1.01 m^2	3.35 m^2
Exit Cross-Sectional Area	4.74 m^2	15.49 m^2
Length of Duct	16.65 m	22.07 m
Field on Axis at Inlet	6.0 T	6.0 T
Field on Axis at Exit	3.5 T	3.5 T
Peak Ampere-Turns Required	2.214×10^7 A-T	3.703×10^7 A-T
Current per Turn	5 kA	5 kA
Average Winding Current Density	1.4×10^8 A/m^2	1.4×10^8 A/m^2
Winding Packing Factor	0.7	0.7
Conductor Aspect Ratio	2:1	2:1
Fraction of Superconductor in Conductor	0.25	0.25
Inductance	161 k	597 k
Stored Energy	2014 MJ	7467 MJ
Number of Turns	4428	7406
Conductor Operating Temperature	4.2°K	4.2°K
Liquid Helium Refrigerator Thermal Load	121 W	277 W
Liquid Helium Refrigerator Electrical Load	163 kW	290 kW
Liquid Nitrogen Refrigerator Electrical Load	71 kW	156 kW
Total Electrical Load	234 kW	446 kW
Total Estimated Cost	$28.2 M	$82.3 M
$/kVA	$47	$41

Table A 9.9.5 Summary of Magnet Design for 2000 MW Fired Cycle (like Base Case 2, Point 1)

<u>Superconductor (Nb-Ti)</u>	
Material Cost \$25/lb @ 2×10^5 A/cm^2 @ 5T	\$14,500,000
Volume, (566 lb/ft^3)[a]	1024
Weight, tons	290
Fabrication, \$16/lb	\$9,300,000
Total Cost	\$23,800,000
<u>Structure (310 SS)</u>	
Material Cost, \$1.60/lb	\$4,960,000
Volume, ft^3	6186
Weight, tons	1550
Fabrication, \$5.40/lb	\$16,740,000
Total Cost	\$21,700,000
<u>Base Cost</u>	\$45,500,000
Engineering	\$11,400,000
Construction	\$11,400,000
Design allowance	\$13,600,000
Total Cost of Magnet	\$81,900,000
Refrigerator Cost	400,000
TOTAL COST	\$82,300,000

[a]Superconductor density.

Table A 9.9.6 Summary of Magnet Design for 600 MW Direct-Fired Cycle (like Base Case 1, Point 3)

Superconductor (Nb-Ti)	
Material Cost $25/lb @ 2×10^5 A/cm^2 @ 5T	$6,380,000
Volume, (566 lb/ft^3)[a]	450
Weight, tons	127.5
Fabrication Cost, $16/lb	$4,080,000
Total Cost	$10,500,000
Structure (310 SS)	
Material Cost, $1.60/lb	$1,150,000
Volume, ft^3	1437
Weight, tons	360
Fabrication Cost, $5.40/lb	$3,890,000
Total Cost	$5,040,000
Base Cost	$15,500,000
Engineering, 25% Base Cost	$3,880,000
Construction, 25% Base Cost	$3,880,000
Design allowance	$4,650,000
Total Cost of Magnet	$27,900,000
Refrigerator Cost	300,000
TOTAL COST	$28,200,000 ± 25%

[a]Superconductor density.

Appendix A 9.10

OPEN-CYCLE MHD GENERATOR CHANNEL

The Base Case 2 MHD generator channel is 22 m (72.2 ft) long with 1.82 m (5.97 ft) square inlet cross section, and 3.91 m (12.83 ft) square outlet. Maximum pressure inside the channel was 405.2 kPa (4 atm) at 2581°K (4186°F). The channel must be designed to contain this pressure at high velocities [775 m/s (2543 ft/s)]. Also, the channel must lie within the field created by the cryogenic magnetic.

A box beam with the dimensions of the MHD channel and with 2.5 cm (0.98/in) thick walls will have a maximum deflection of 0.02 cm (0.78 in) when supported at the ends only. However, an unstiffened 2.5 cm (0.984 in) thick wall will not contain the pressure load. In Base Case 2 heat loss through the wall of the MHD channel is used to heat the boiler feedwater. When the pipes carrying the boiler feedwater are used to strengthen the channel walls, a 2.5 cm (0.984 in) thick wall is adequate.

Figure A 9.10.1 shows the relation of the MHD generator channel to the magnet and dewar. Figure A 9.10.2 shows the arrangement of the boiler feedwater pipes around the channel.

The interior of the MHD channel will be insulated with magnesium oxide (MgO) blocks to protect the inconel walls from the high-temperature MHD gas. The insulating blocks will be 10 by 10 by 2 cm (3.94 by 3.94 by 0.787 in.) thick and will be installed as a mosaic. The small size is necessary to prevent thermal cracking. The magnesium oxide will be mounted to a 0.5 m (19.68 in) square mounting plate as shown in Figure A 9.10.3. The mounting plate will be attached to the channel walls with studs. Transpiration air will infiltrate the gas stream from between the blocks. This will form a laminar layer over the face of the blocks and reduce the block temperature.

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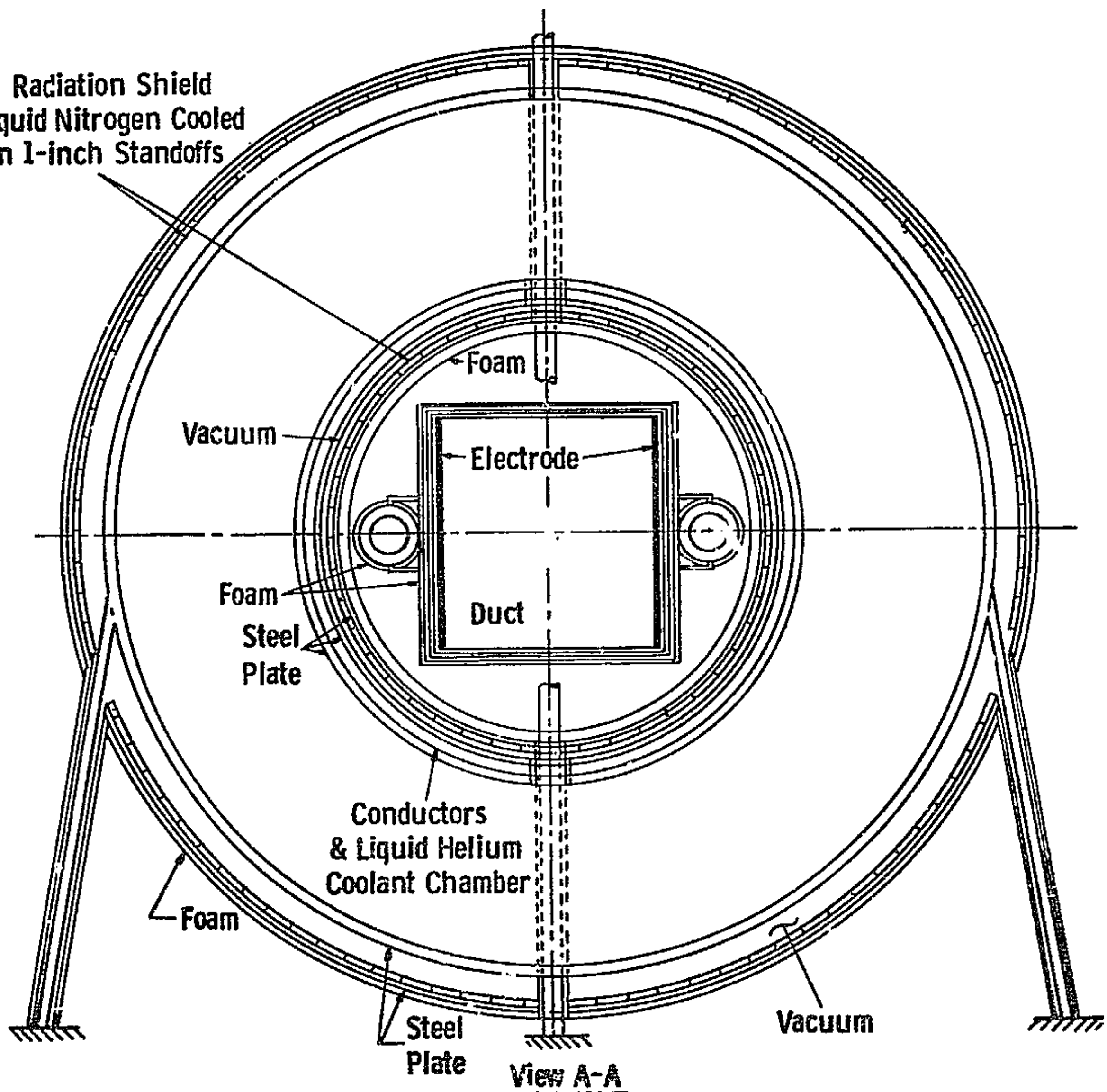

Fig. A 9.10.1—Cross sectional view of MHD duct and magnet enclosure

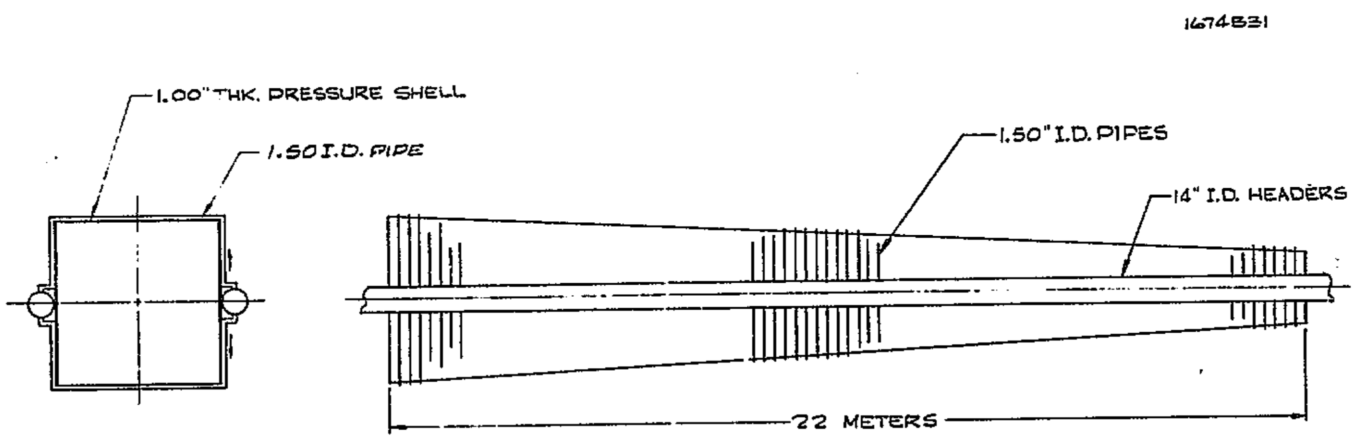

Fig. A 9. 10. 2—Base Case 2 duct and boiler feed water pipes

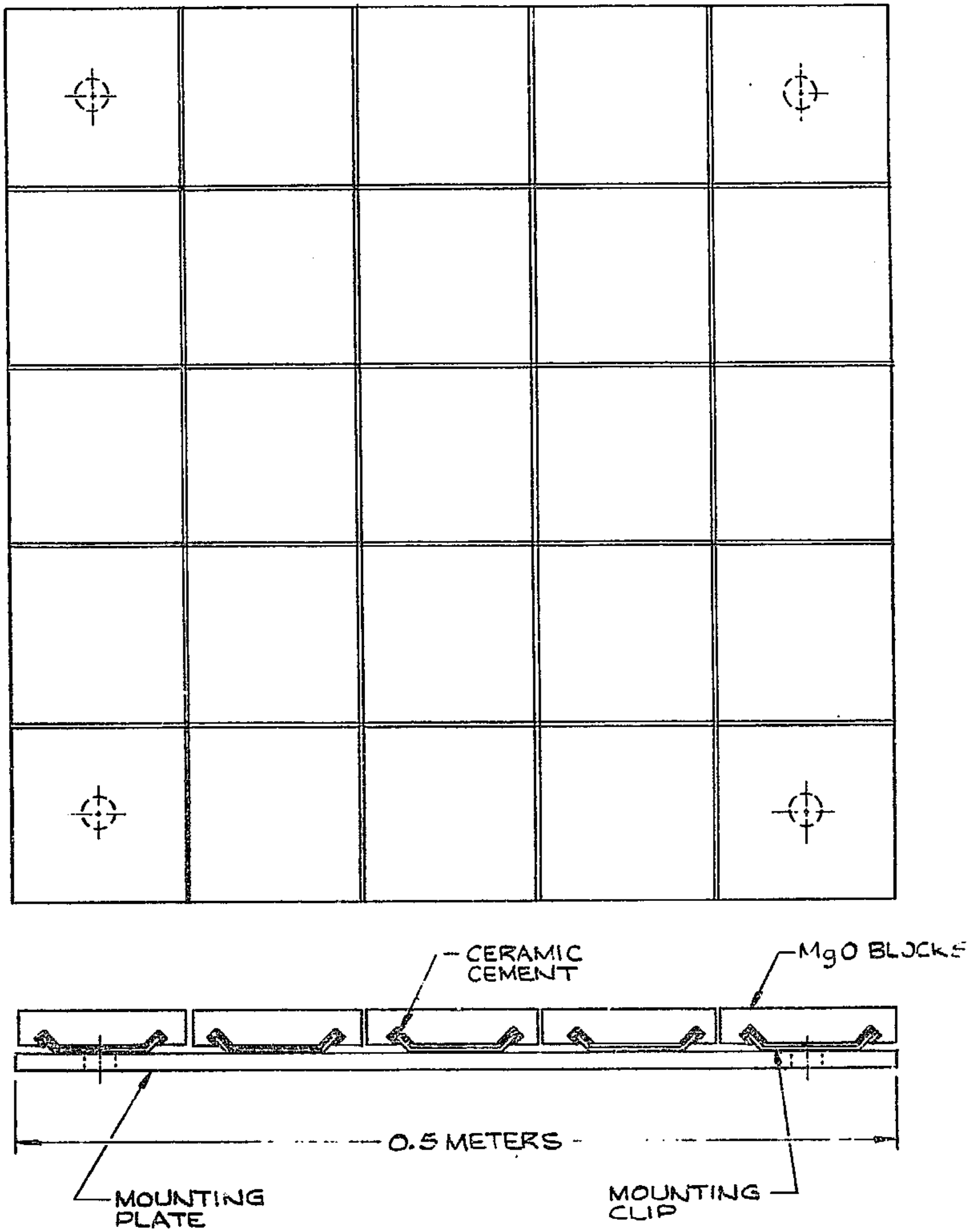

Fig. A 9.10.3—Duct insulation module

The electrodes will be of the same mosaic construction but of different ceramic materials. Electrical connections will be made to the appropriate mounting plates to give the desired electrical configuration.

An estimate of the weight and cost of the generator channel is given in Table A 9.10.1.

Table A 9.10.1 Weight and Cost of Generator Dect for Base Case 2

	Weight, lb	Material Cost, $
Boiler Feedwater Headers	136,783	360,000
Boiler Feedwater Heater Tubes	15,427	45,000
Pressure Shell	108,626	325,878
Side-Wall Insulation	18,600	55,800
Electrode Wall Ceramics	18,600	55,800
Insulation Support Plates	27,589	82,767
Transpiration Air Piping	381	381
Electrical Connectors	1,000	5,000
TOTAL	327,006	885,059
	Installation Labor	$1,770,118
	Total Cost	$2,655,118

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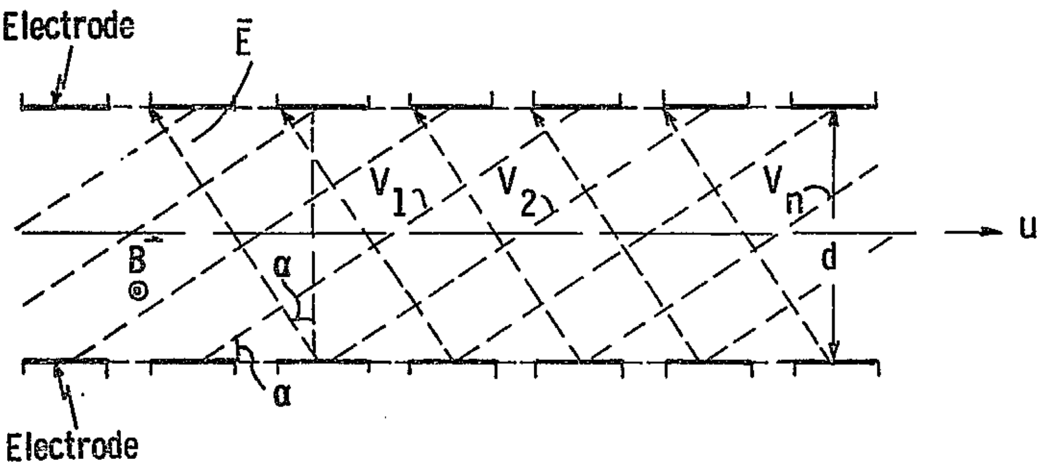

Fig. A 9. 11. 1—Electric field plot of a typical segmented electrode MHD duct. $\bar{E}$—the internal electric field vectors, V_n—the nth equipotential line, u—gas velocity, B—magnetic flux density, and $\alpha = \arctan\,(1-K)\beta/K$ where K—generator coefficient and β—hall coefficient

Appendix A 9.11

SIZE, WEIGHT, AND COSTS OF DC TO AC POWER CONDITIONING SYSTEM FOR OPEN-CYCLE MHD GENERATORS

The size, weight, and costs of dc-to-ac power conditioning systems to be used with MHD generators were analyzed. Since MHD generators are dc devices, their dc power has to be converted to ac to be compatible with the steam plant and for power transmission. The dc-to-ac converters can be either static-type (solid state) or mechanical-type (motor-generator sets). For the open-cycle MHD generator in which the output voltage is 14 kV or higher, output currents of 2-5kA are possible, and solid state converter is plausible. The efficiency of such a system is on the order of 98 to 99% which represents a very low power loss.

The proposed open-cycle MHD generator duct contains segmented electrodes. Since the working fluid is combustion gas with potassium or cesium seed, the electrical conductivity is low compared to liquid-metal MHD, and the Hall currents are low. Subsequently, the transverse currents dominate the operating characteristics of the duct. Figure A 9.11.1 shows the internal electric field vectors and the equipotential lines in such a duct. The electric field has a transverse component, E_y, and an axial component, E_x. The angle α between the electric field vector and E_y is related to the Hall angle θ and generator coefficient, K. Since equipotential lines are normal to the electric field, it can be shown that the angle between the equipotential lines and the downstream axis of the duct is also α. When electrodes are connected in series, therefore, one electrode must be connected to the diagonally opposite electrode and the angle of the diagonal with respect to the axis is α. In this manner, pairs of electrodes can be connected in series to generate output voltages compatible with solid-state converters. The width of the electrodes are selected so that the current per electrode is the same.

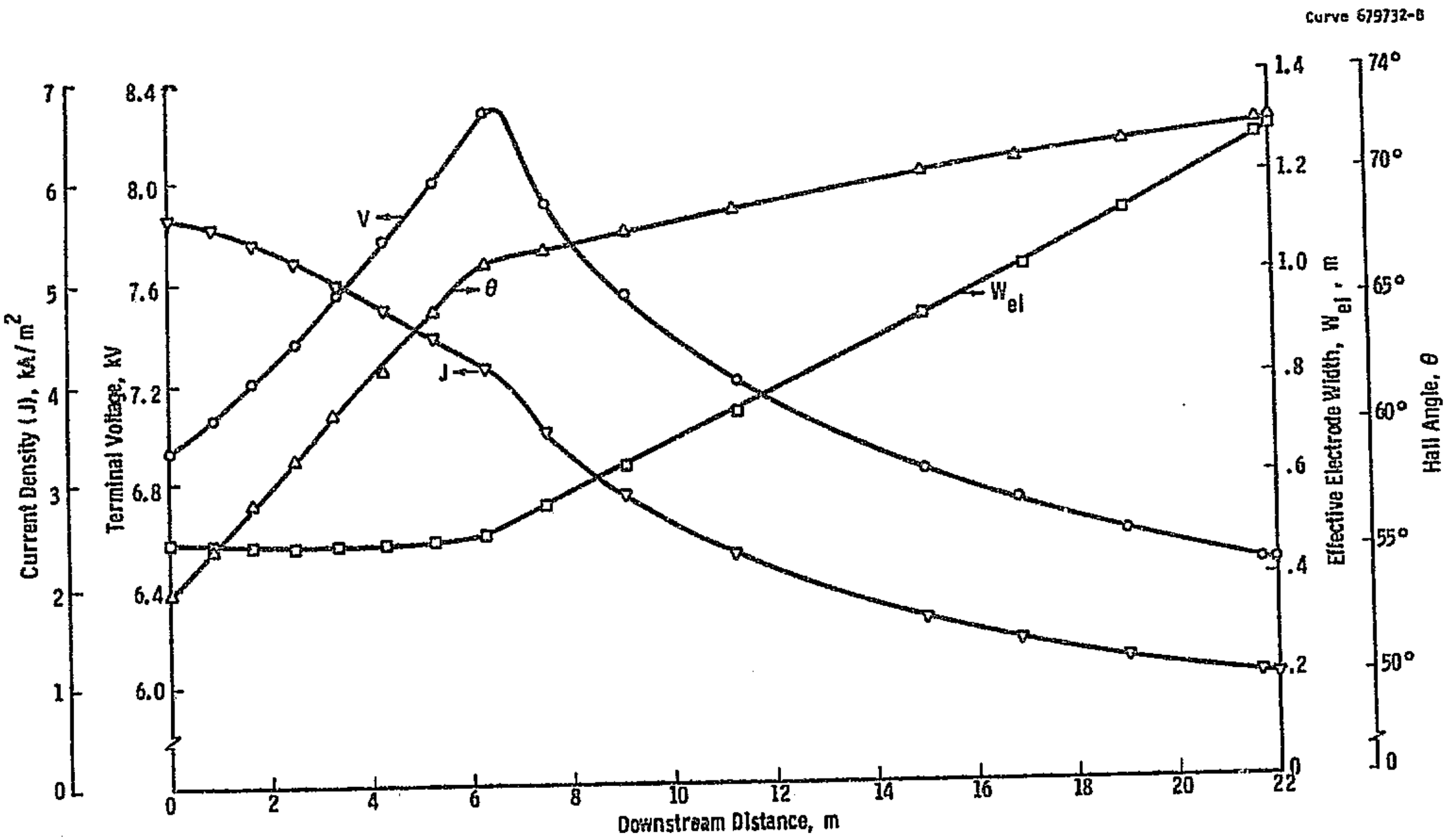

Fig. A 9.11.2 —Electrical characteristics of MHD duct for Base Case 2

A 9.11.1 Detail Example - Base Case 2, Point 1

Figure A 9.11.2 shows the electrical characteristics of the MHD generator as a function of downstream distance. This figure gives the terminal voltage, current density, Hall angle, and effective electrode width, assuming a 10 kA electrode current. The elements making up each electrode were assumed to have been paralleled into two separate groups, each of which would carry 5 kA and be connected to an inverter bank (5 kA being the maximum current rating of present-day systems). The values of current density in Figure A 9.11.2 were used together with Equation A 9.11.1 to determine the electrode widths in the axial direction at 18 positions along the length.

$$W_{el} = \frac{10000}{J\ L} \qquad \text{(A 9.11.1)}$$

where W_{eb} is the electrode width in the axial direction

L is the MHD duct width

J is the current density and the electrode surface

Figure A 9.11.3 shows a typical layout of the electrodes. The diagonal dashed lines indicate equipotential lines between diagonally opposite electrodes. The angle between these dashed lines and the downstream axis varies with downstream distance according to the Hall coefficient, β, times (1-K)/K where K is the generator coefficient. Voltages between pairs of electrodes are given. Note that only three sets of electrode circuits are shown. Two electrode pairs are in series which would generate 14 kV. A total of 17 circuits are required to convert the 1180 MW, dc output power to ac power. A more detailed study of electrode sizes and electrode arrangements is needed to obtain the optimum number of circuits. The arrangement given here, however, is good enough to determine weight, size, and cost of the dc-to-ac converters and associated hardware.

A 9.11.2 D.C. to A.C. for Base Case 2, Point 1

Each 14 kV, 5 kA dc-to-ac converter circuit consists basically of two parallel inverters having the outputs connected to primaries of the

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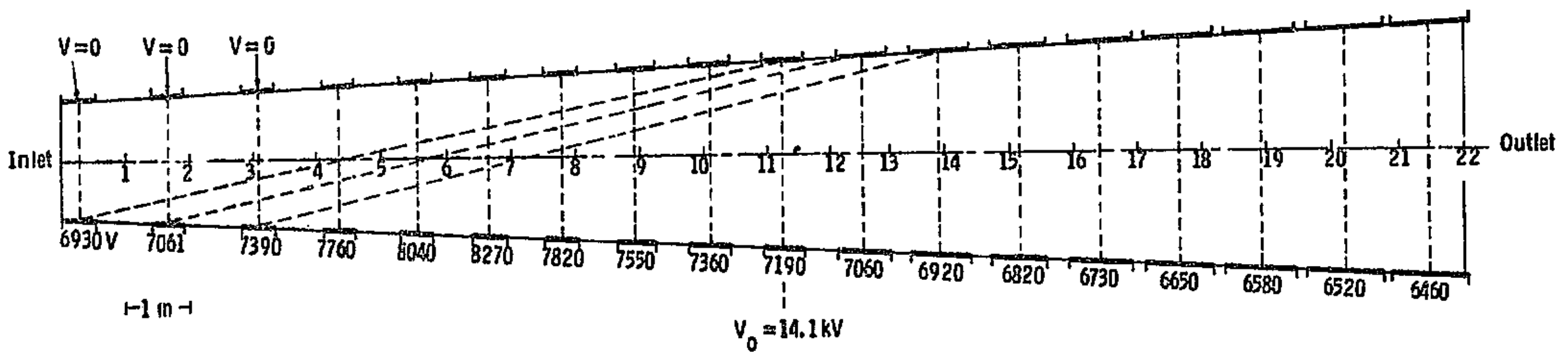

Fig. A 9.11.3—Layout of open-cycle MHD duct—Base Case 2. Shown is typical arrangement for connecting electrodes in series. The heavy lines are electrodes, diagonal dash lines are equipotential lines, V_0 is output voltage for two electrodes in series. Assumed current per electrode is 5 kA.

power transformers. Figure A 9.11.4 shows a typical circuit of a 12-pulse converter. The dc current from the MHD generator electrode circuit is fed (1) through a dc filter reactor used as a smoothing choke, (2) to an interphase reactor, (3) from each leg of the interphase reactor to a 6-pulse thyristor bridge inverter, converting the dc to three-phase ac pulses. The output of one thyristor bridge is fed to a delta-connected primary winding of the inverter transformer and the other thyristor bridge is fed to a wye-connected primary winding of the same transformer. The 34.5 kV secondary delta-connected winding of each inverter transformer is fed through an ac circuit breaker and to the primary of the power transformer. The secondary voltage of the power transformer is assumed to be 500 kV. Actually, this voltage may be higher. Since the inverter is a 12-pulse scheme, a series of harmonics of order 11, 13, 23, 25, 35, 37.... and amplitudes of 1/11, 1/13.... are generated and fed into the line. A tunable filter (not shown in Figure A 9.11.4, therefore, is connected across the line to filter out the harmonics. Disconnect switches are placed between the inverters and the transformers.

Figure A 9.11.5 shows a block diagram of the dc interrupter, the inverters, disconnect switches, transformers, ac circuit breakers, line filter, and tie breaker of the overall power conditioning system.

A 9.11.3 Size, Weight and Cost of Power Conditioning System for Base Case 2, Point 1

A detail analysis of the size, weight, and cost of a power conditioning system for a 26 MW fuel cell has been made by the Westinghouse Electronic Department. The size and weight of the components for the system for Base Case 2 (1180 MW) were scaled from this 26 MW system. Table 9.11.1 shows a breakdown of size and weight of the 26 MW system. The various items can be identified with the components shown in Figure A 9.11.4. The scaling factor used was the ratio of the MWs raised to a power. This exponent can vary from 0.75 to 1. We chose an intermediate scaling exponent of 0.85. The size and weight of the inverters were calculated by knowing the number of thyristors to be used and from previous experience in building such devices (Table A 9.11.2). The cooling system for the

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Inverter Scheme

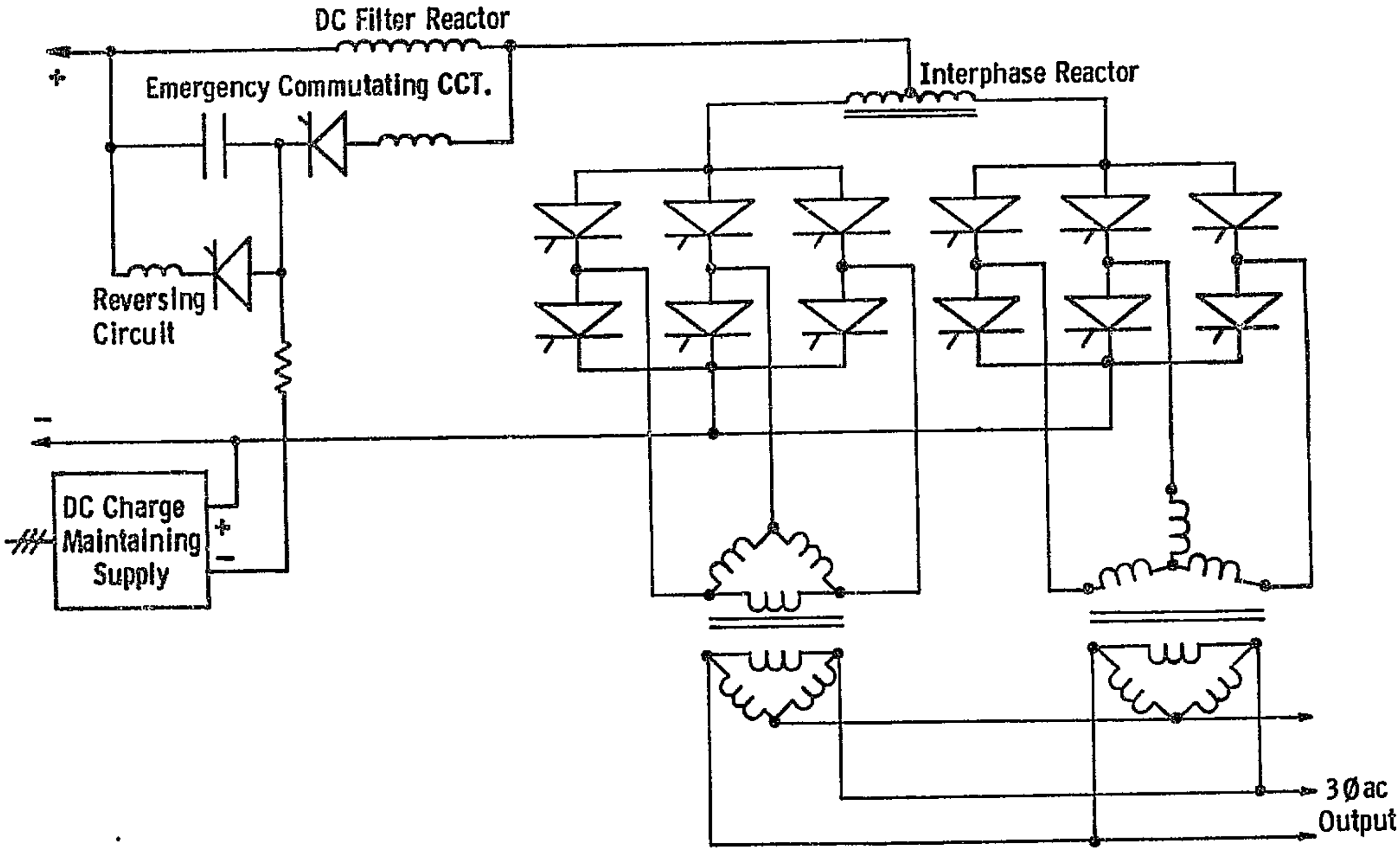

Fig. A 9.11.4 – Inverter scheme for dc to 3Ø ac power conversion

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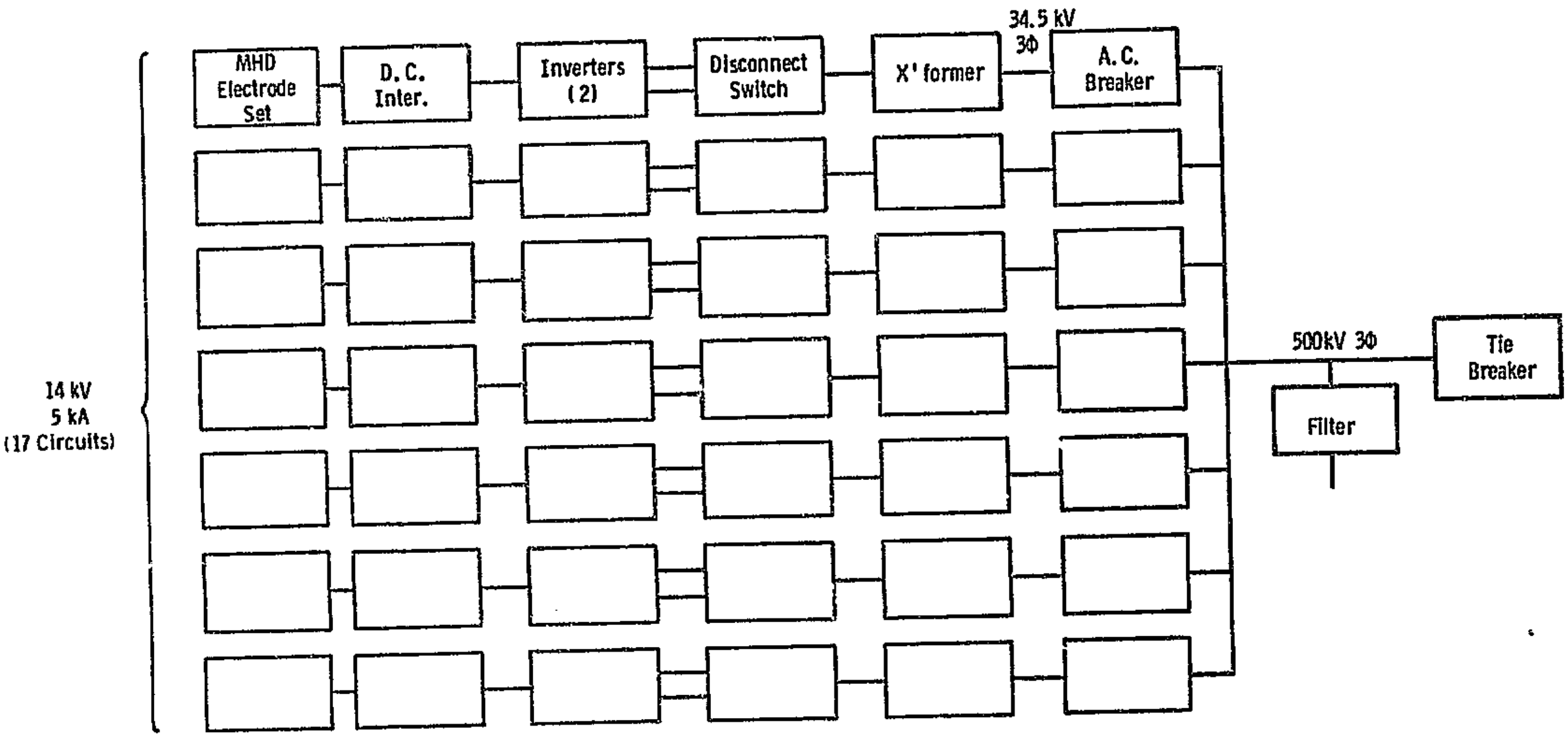

Fig. A 9.11.5—Block diagram of dc to ac power condition system for Base Case 2

Table A 9.11.1 Inverter Scheme for 26 MW

Item	Dimensions, in	Weight, lb
Filter C, 11th Harmonic	78 x 180 x 36	12,000
Filter L, 11th Harmonic	115 x 32	2,300
Filter C, 13th Harmonic	78 x 130 x 36	8,400
Filter L, 13th Harmonic	115 x 32	2,300
Filter C, High Pass	78 x 90 x 36	4,500
Filter L, High Pass	122 x 36	2,100
PF Correction C	78 x 144 x 90	25,000
PF Correction L	No data available	
Transformers, 2 each	No data	Est. 111,000
Inverter Interphase Reactors, 4 each	75 x 37 x 43	2,950
DC Filter Reactors, 4 each	139 x 46	5,400
Emergency Commutating Capacitors, 4 each	36 x 54 x 40	3,600
Emergency Commutating di/dt Reactor, 4 each	23 x 36	600
Emergency Commutating Reversing Reactor, 4 each	29 x 25	275

Table A 9.11.2 1180 MW System

A.70 MW Direct Current - Alternating Current Converter Circuit

Item	Dimensions, in	Weight, lb
Inverter Set	10H x 13L x 7w	13,650
Inverter Cooling System	8 x 10 x 6	20,475
Inverter Interphase Reactor	12 x 6 x 7	22,240
DC Filter Reactor	22.7H x 7.5D	40,711
Emergency Commutating Capacitors	6.6 x 9.9 x 7.3	38,772
Emergency Commutating di/dt Reactor	3.7 x 5.9	4,523
Emergency Commutating Reversing Reactor	4.7 x 4.1	2,073
Inverter Transformer	28 x 15 x 16	180,000
AC Circuit Breaker	11 x 5.1 x 13.4	9,620
TOTAL Converter Circuit Weight		332,064

B. 1100MW Filter

Filter C, 11th Harmonic	22.6 x 52.3 x 10	
Filter L, 11th Harmonic	27 x 7.5	
Filter C, 13th Harmonic	22.6 x 37.7 x 10	734,800
Filter L, 13th Harmonic	27 x 7.5	
Filter C, High Pass	22.6 x 26.1 x 10	
Filter L, High Pass	29.4 x 8.6	
Power Transformers	38.7 x 21.7 x 17.5	1,335,000
TOTAL Weight of System		7,714,900

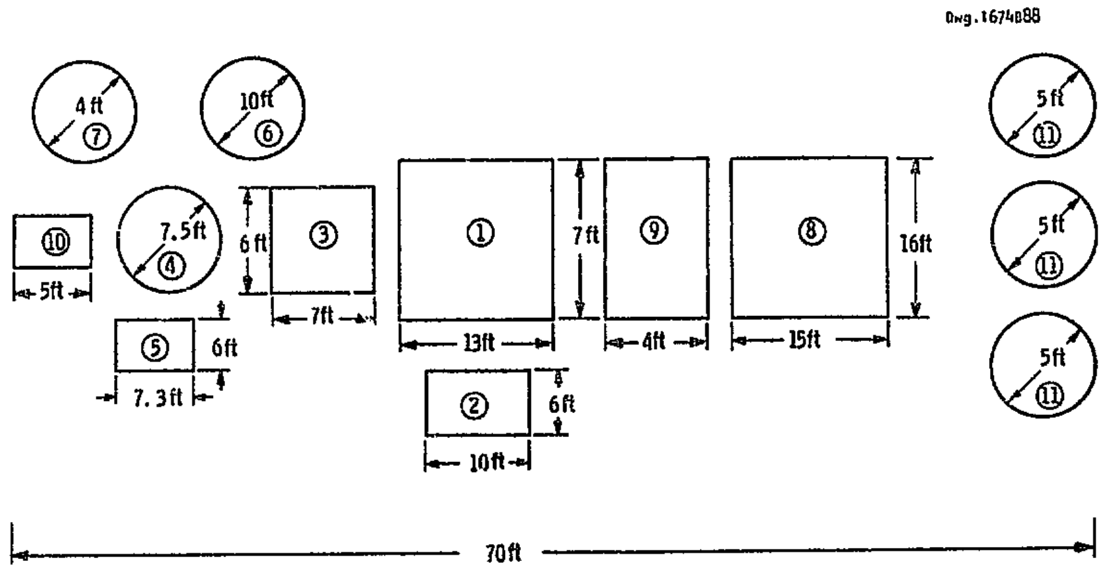

Fig. A 9.11.6 — 70 MW dc - ac converter circuit layout. See Table A 9.11.4 for component identification

thyristors is approximately one-third the volume of the thyristor stacks, stocks and the weight for the 70 MW 12-pulse inverter is 9299 kg (20,500 lb), also based on experience. The sizes and weights of the transformers were obtained from data supplied by the Westinghouse Transformer Division. The pad required to house the power conditioning system is estimated to be 65 m (196 ft) long and 270 m (680 ft) wide. The overall weight of the components is 3,499,000 kg (7,714,900 lb). Layouts of the 70 MW circuit components and filter circuit components are shown in Figure A 9.11.6 and A 9.11.7. Table A9.11.3 explains the numbers in the preceeding figures.

A 9.11.4 Cost of Power Conditioning System for Base Case 2

The cost of a 1180 MW dc-to-ac power conditioning system for Base Case 2 is based on the following cost/kW for the dc components or cost/kVA for the ac components according to the following schedule:

- Inverter (including dc filter reactor, interphase reactor, commutating circuits, inverter bridges, and cooling system) - $30/kW
- Inverter transformers - $7.2/kVA
- Tunable filter - $6/kVA
- dc interrupter - $6/kW
- Power transformer - $2.5/kVA

Results of cost calculations are given in Table A 9.11.4. The total cost of the 1180 MW system is $57.8 million, which represents a unit cost of approximately $49/kW. These costs are FOB and do not include delivery or installation. The Power System Planning group at Westinghouse was consulted about the costing of this dc-to-ac power conversion system. They arrived at a cost range of $40 to 50/kW. The most probable cost of this power conversion systems, therefore, is assumed to be $50/kW. This cost is lower than that estimated for the smaller 26 MW system, which was $58/kW. An estimate of the installation costs for the inverter systems and tuned filter is given in Figure A 9.11.8.

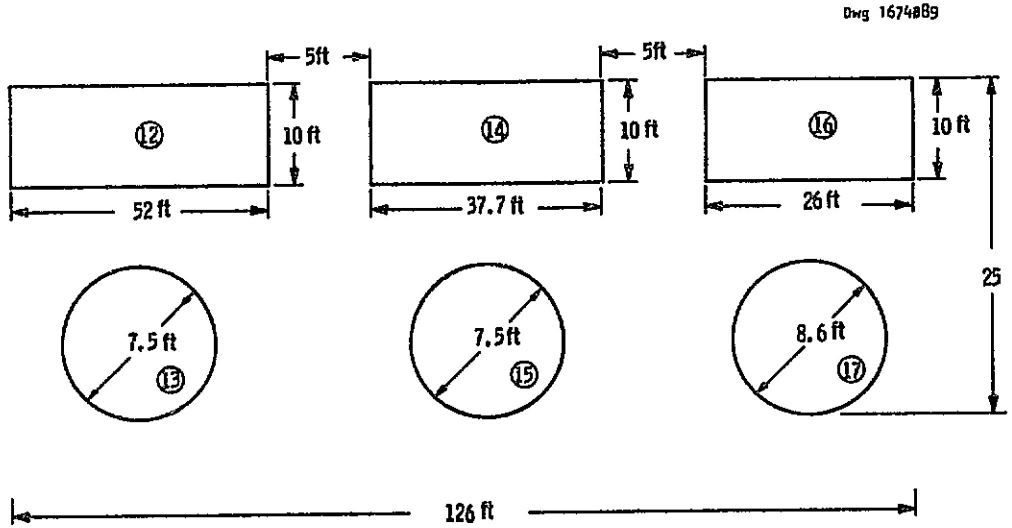

Fig. A 9.11.7 — Layout of filter for 1180 MW system. See Table A 9.11.4 for component identification

Table A 9.11.3

Items shown in Layout of Converter Circuits (Figures A 9.11.6 and A 9.11.7)

Item No.	Item
1	Inverters
2	Inverter Cooling System
3	Inverter Interphase Reactor
4	DC Filter Reactor
5	Emergency Commutating Capacitors
6	Emergency Commutating di/dt Reactor
7	Emergency Commutating Reversing Reactor
8	Inverter Transformer
9	Disconnect Switch Array
10	DC Interrupter
11	Lightning arresters
12	Filter C, 11th Harmonic
13	Filter L, 11th Harmonic
14	Filter C, 13th Harmonic
15	Filter L, 13th Harmonic
16	Filter C, High Pass
17	Filter L, High Pass

Table A 9.11.4 Cost Breakdown of Base Case 2 (1180 MW)

A. 14 KV, 5 kA dc Circuit		
Inverters	$ 2,100,000	
Inverter transformer	504,000	
DC interrupter	420,000	
AC circuit breaker	24,200	
TOTAL	$ 3,048,200	
TOTAL for 17 circuits		$51,819,400
B. Filter Circuit and Power Transformer		
Filter	$ 3,000,000	
Power transformers	2,950,000	
Total		5,950,000
Total System Cost		$ 57,769,400
Cost, $/kW		$48.9/kW

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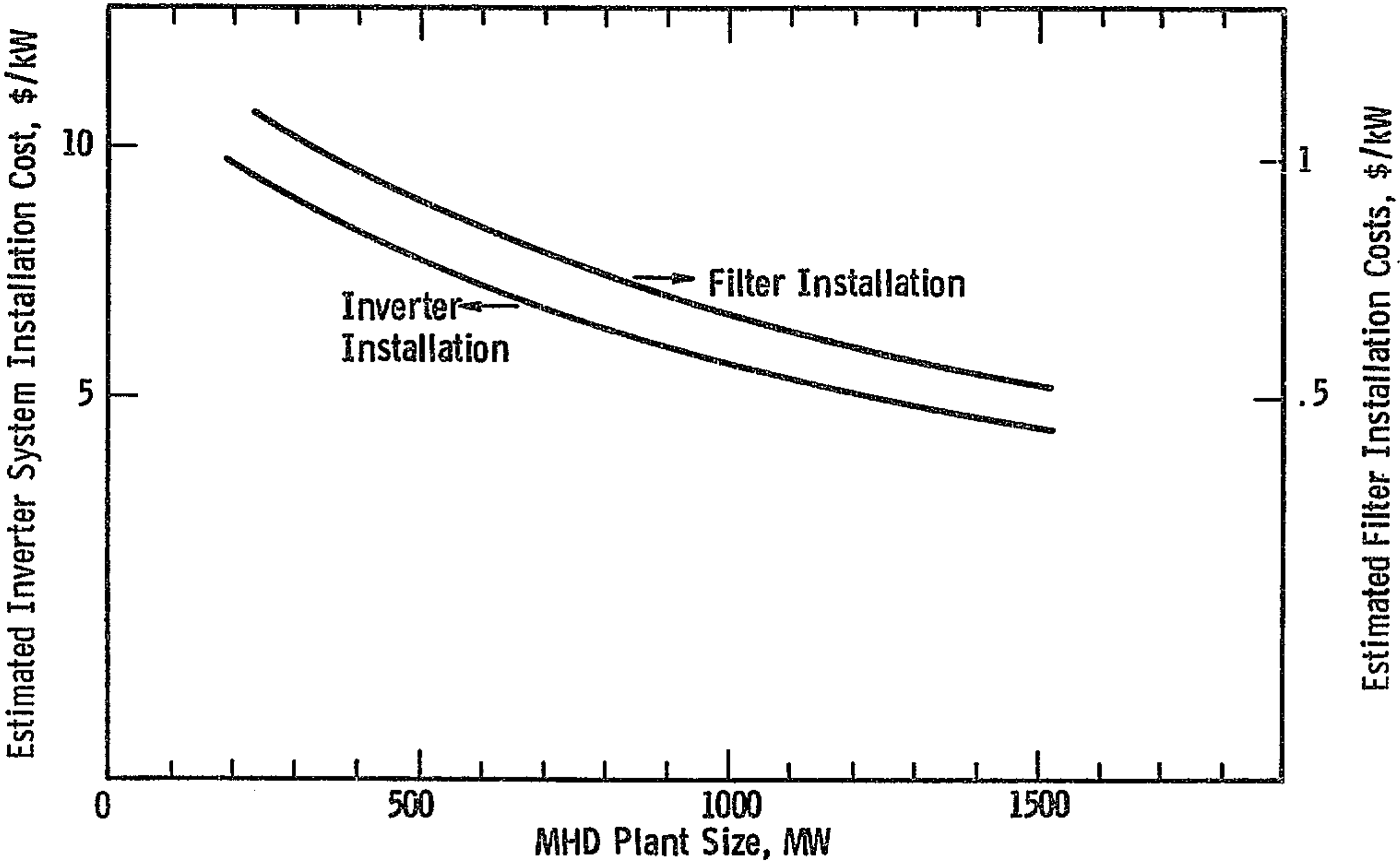

Fig. A 9. 11. 8—Typical estimates of installation costs for inverters and tuned filters versus MHD plant size

A 9.11.5 Circuit Protection

A 9.11.5.1 Short Circuits

When a short circuit occurs in the MHD generator, the thyristor bridges should prevent a back flow of current from the other circuits of the MHD generator or from the main transmission lines. If, on the other hand, the thyristors should become shorted, the dc breaker between the MHD generator electrodes and thyristor bridge should interrupt the current and isolate that circuit from the system.

When a short occurs on the transmission line, the MHD generator will deliver a short circuit current of approximately five times the normal load current. This short circuit will be interrupted either by the dc interrupter or by the tie-line breaker. Fortunately, an MHD generator is a rather "soft" current source compared to an ac alternator.

A 9.11.5.2 Voltage Spikes

Voltage instabilities or oscillations could develop in the MHD generator, or the load current may suddenly go to zero. The thyristor inverter bridge is designed to handle these transients without "blowing out" components. For example, the 14 kV thyristor bridge, transformers, and reactors have a 42 kV transient voltage rating known as the BIL (Basic Impulse Level). Since some of the electrodes in the MHD duct are connected in series, voltage instabilities in these circuits should average out. In addition, the dc filter reactor should help to isolate the thyristor bridge from voltage spikes generated in the MHD duct. The dc reactor is designed to smooth out voltage oscillations of frequencies greater than 1 kHz.

A 9.11.5.3 Lightning Strokes and Switching Transients

The transmission line which is connected to the output of the power conditioning system is subject to lightning and switching surges. These transients are reflected back to the inverter transformers. The transformers are designed to handle such transients without being damaged. Besides, normal practice is to protect the transformers by placing lightning

arresters from line to ground at the transformer sites. These lightning arresters are designed to handle any power follow current that flows after being activated.

A 9.11.5.4 System Stability to Load Changes

The MHD power plant supplies power from both the MHD generator and the ac generator of the steam bottoming plant. These two generators working together have a stabilizing effect on the system voltage and frequency when load changes or when one of the dc to ac inverters is suddenly disconnected. Lips and Ring (Reference 9.43) have made measurements on a similar system in which the dc to ac inverters supply 500-1000 MW and the ac generator supplies 140-280 MVA. Their experiments included switching the inverters in and out of the system and switching the ac generator. The results of the experiment show:

- Connection of high voltage dc inverters to different buses of a common ac system does not lead to increased voltage distortion.
- Alternating current voltage can be easily controlled by a combined action of generator excitation and controllable static compensators including switch capacitors, even if the total inverter rating throughout the system exceeds the installed synchronous generator capacity.
- Alternating current systems with predominant dc injection by the inverters can be controlled to work stably under all normal and most fault conditions.
- Successful operation of a conventional line commutated high-voltage dc inverter supplying a purely passive load shows that there is no technizal limit to the amount of dc injection by inverters into a given ac system.

Lips (Reference 9.44) discusses the effect of multiple in feed of high-voltage dc inverter stations into a common ac system. He shows how these inverters can inject harmonics in the system and cause instabilities in the system.

Appendix A 9.12

SEED COLLECTION

Three broad options exist for the collection of seed compounds and fly ash from the open-cycle MHD system;

- Wet scrubbing
- Fabric filtration
- Electrostatic precipitation

Wet scrubbing offers one potential advantage. If seed reprocessing is to incorporate aqueous treatment, the scrubber will perform dual functions - collecting the seed and simultaneously dissolving the soluble potassium or cedium salts. If dry processing of the seed is anticipated, however, dissolution of the seed material complicates the process and requires evaporation and recrystallization steps.

As dry processing of the seed is projected, wet scrubbing is not considered appropriate for seed collection.

Bag filters have been used successfully on a laboratory scale for the collection of MHD seed material, (Reference 9.4 and 9.45). The experience of the English program (Reference 9.4) showed a tendency towards bag "blinding", with associated problems of bag cleaning and excessive operating pressure drop. Experience with other dusts which blind bag filters (Reference 9.45) indicates that the necessarily vigorous bag cleaning techniques result in accelerated bag failure rates, loss of material, and continuous maintenance difficulties.

Electrostatic precipitation offers the potential for meeting seed recovery and emission requirements, shows no unusual operating difficulties (at this stage), and is closely allied to current power plant practice.

Based on the above considerations, we have selected electrostatic precipitators for seed collection in the open-cycle MHD systems which are under evaluation.

A 9.12.1 Preliminary Design Estimates

Precipitator sizing is based on laboratory data established in the English MHD program (Reference 9.4).

A 9.12.1.1 Flow Cross Section

Total gas flow to the precipitators will be approximately 1588 kgls (3500 lb/s) or 1981 m^3/s (70,000 ft^3/s) Based on the English data (Reference 9.4) an optimum gas velocity through the precipitator is 1.37 m/s (4.5 ft/s). This requires a flow area of 1440 m^2 (15,500 ft^2). Assuming an active precipitator height of 15.24 m (50 ft), the precipitator width will be 96.0 m (315 ft). Alternatively, two units 15.2 m by 48.8 m (50 ft by 160 ft) could be used (Figure 9.12.1).

A 9.12.2.1 Plate Area

Plate area can be estimated from the relationship:

$$\eta = 1-e^{-(CA/Q)} \qquad (1)$$

where

η is the required precipitator efficiency

C is the effective migration velocity of particles in the precipitator

A is the precipitator plate area

Q is the gas flow rate

For Base Case 2, a design efficiency of 99.5% is required to comply with present particulate standards. For seed precipitation under acceptable precipitator operating conditions, the effective migration velocity is .0427 m/s (0.14 ft/s). This requires a collection plate area of 0.269 $(km)^2$. (2.9×10^6 ft^2).

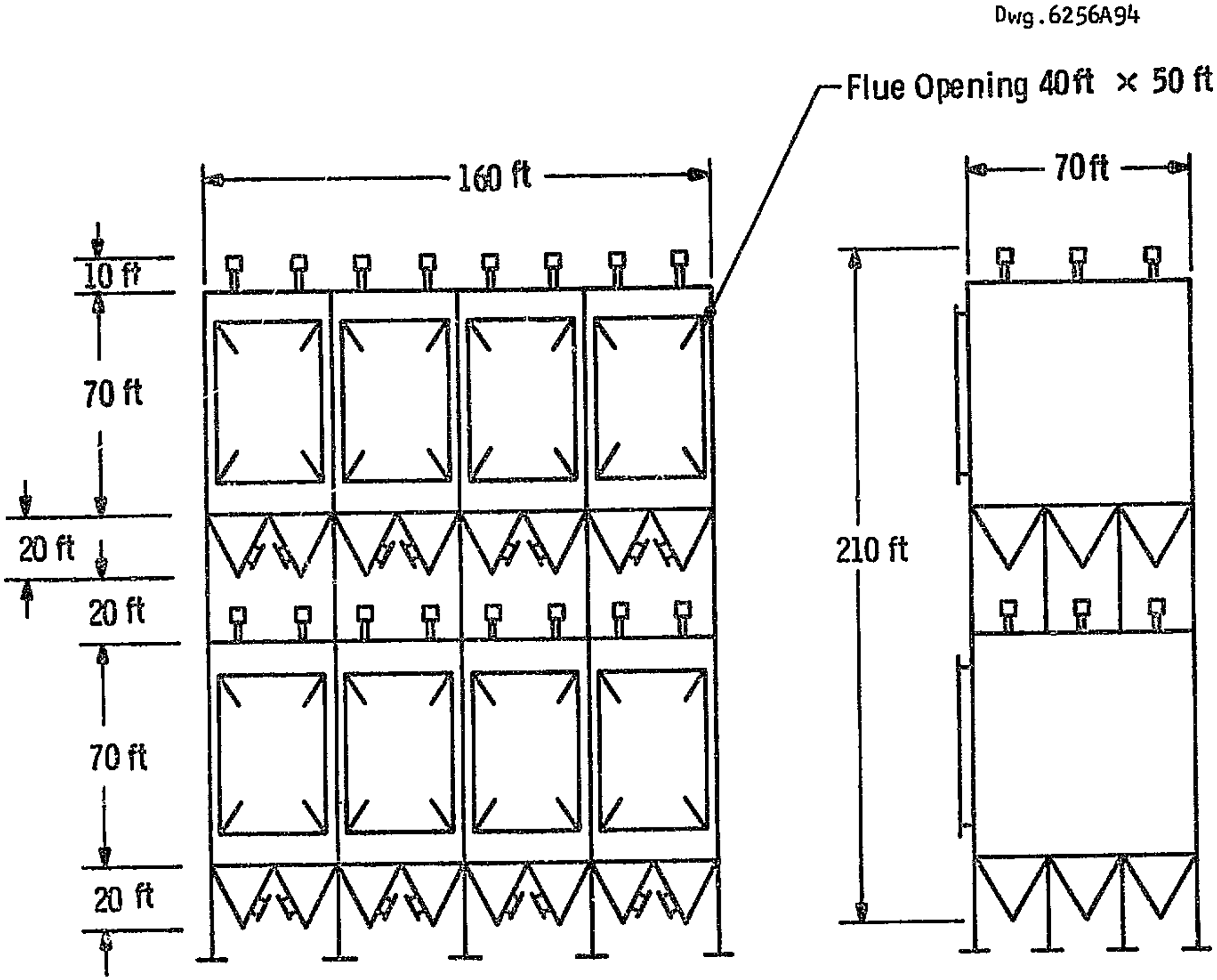

Fig. A 9.12.1 —Electrostatic precipitator layout

Assuming an 0.203 m (8 in) spacing for precipitator plates, 0.269 (2.9×10^6 ft^2) of collection surface can be accommodated in a precipitator 15.24m (50 ft) high by 96.m (315 ft) wide and 21.3 m (70 ft) deep, or in two units 15.24 m (50 ft) by 48.77 m (160 ft) by (70 ft) (Figure A 9.12.1).

Equation A 9.12.1 was used to estimate the required precipitator plate area for other open-cylce MHD parametric points. Capital costs for these cases were scaled from the base case using the relationship.

$$\frac{\$}{\$B} = \left(\frac{A}{A_b}\right)^{0.8}$$

where $ is the capital cost of the precipitator

$B is the capital cost of the base case precipitate

A is the plate area of the precipitator

A_b is the plate area of the base case precipitator

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